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Union Carbide Corp, Tonawanda, NY

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Biological-Chemical Process for Removing Phosphorus at Reno/Sparks, NV

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16. ABSTRACT
This demonstration project was initiated with the purpose of establishing the capability of the PhoStrip system to remove phosphorus on a full scale activated sludge system and compare these results to those obtained on a pilot scale system. This process employs the capability of activated sludge to take up and release phosphorus as a result of being cycled between aerobic and anoxic conditions. During the aerobic period in the aeration basin the phosphorus in the wastewater is removed by the biomass. The sludge is then separated from the phosphorus-free wastewater in the secondary clarifier. A portion of the return sludge from the clarifier is pumped to an anoxic stripper tank where the phosphorus is released. This phosphorus-containing water is then separated from the sludge either through further thickening of the sludge or through elutriation. The supernatant from the anoxic zone is treated with lime to precipitate the phosphorus and the sludge from the anoxic zone is returned to the aeration basin. The pilot and full scale testing was performed at the Reno/Sparks Joint Water Pollution Control Plant located in Sparks, Nevada. The pilot scale system consisted of a 0.11 m^3/hr diffused air activated sludge system. This system was designed to produce optimum performance with the elutriation mode of separating the phosphorus from the sludge in the anoxic zone (stripper tank). The full scale system consisted of a temporary modification of the Reno/Sparks plant. Treating one-third of the plant flow (25,000 m^3/d) was first tested using the thickening mode of separating phosphorus from the sludge in the thickener. Later two-thirds of the plant flow (51,000 m^3/d) was treated utilizing the elutriation method.

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BIOLOGICAL-CHEMICAL PROCESS FOR REMOVING PHOSPHORUS AT RENO/SPARKS, NV

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

This report describes those features of process theory, plant design, and operational control necessary to remove phosphorus from an activated sludge system employing a biological-chemical process.

Francis T. Mayo, Director
Municipal Environmental Research
Laboratory

ABSTRACT

This demonstration project was initiated with the purpose of establishing the capability of the PhoStrip system to remove phosphorus on a full scale activated sludge system and compare these results to those obtained on a pilot scale system.

This process employs the capability of activated sludge microorganisms to take up and release phosphorus as a result of being cycled between aerobic and anoxic conditions. During the aerobic period in the aeration basin the phosphorus in the wastewater is removed by the biomass. The sludge is then separated from the phosphorus-free wastewater in the secondary clarifier. A portion of the return sludge from the clarifier is pumped to an anoxic stripper tank where the phosphorus is released to the water surrounding the sludge. This phosphorus-containing water is then separated from the sludge either through further thickening of the sludge or through elutriation. The supernatant from the anoxic zone is treated with lime to precipitate the phosphorus and the sludge from the anoxic zone is returned to the aeration basin.

The pilot and full scale testing was performed at the Reno/Sparks Joint Water Pollution Control Plant located in Sparks, Nevada. The pilot scale system consisted of a 0.11 m^3/hr. diffused air activated sludge system designed to incorporate the PhoStrip system. This system was designed to produce optimum performance with the elutriation mode of separating the phosphorus from the sludge in the anoxic zone (stripper tank). The full scale system consisted of a temporary modification of the Reno/Sparks plant. Treating one-third of the plant flow (25,000 m^3/d) was first tested using the thickening mode of separating phosphorus from the sludge in the thickener. Later, two-thirds of the plant flow (51,000 m^3/d) was treated utilizing the elutriation method.

The PhoStrip system demonstrated the capability of reducing the effluent total phosphorus to less than 1 mg/l on both the pilot and full scale systems. The factors involved in designing the process to achieve these levels were discussed and recommendations were made as to the proper selection of conditions to achieve the desired performance. A method of predicting lime requirements and chemical sludge production was proposed. The overall cost of the system relative to conventional mineral addition systems was discussed.

This report was submitted in fulfillment of EPA Grant No. R804931-01 by the City of Sparks, Nevada under the sponsorship of the U.S. Environmental Protection Agency. This report covers the period October 26, 1976 to January 30, 1977, and work was completed January 30, 1977. The report was prepared by Union Carbide Corporation under subcontract to Sparks, NV.

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We would like to thank Mr. Robert Churn of the City of Sparks, Nevada and the Reno/Sparks Joint Committee for permitting us to use their fine facilities and their invaluable aid in implementing the modifications required to convert their plant into a PhoStrip system.

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SECTION 1

INTRODUCTION

A variety of methods have been developed to remove phosphorus from wastewaters. Typically, phosphorus removal is achieved by precipitating phosphates with salts of iron, aluminum, and calcium (lime). The EPA Process Design Manual for Phosphorus Removal (1) indicates that precipitation can be accomplished in the primary clarifier, as part of the secondary treatment process, or as a tertiary treatment process. The use of lime is generally limited to primary clarifier addition or tertiary treatment because precipitation with lime is pH dependent and secondary systems, i.e., activated sludge, do not perform well at the pH range best suited to the use of lime ($9 \leq pH \leq 12$).

The capability of the activated sludge process to remove substantial quantities of phosphorus from wastewaters without the aid of direct chemical addition is a fairly recent discovery. However, the capacity of microorganisms to accumulate phosphorus is not a new observation. Volutin (a metachromatic granule composed of polyphosphates) was first observed by Ernst (2) in 1888. Since that time, microorganisms such as bacteria, fungi, yeasts, algae and protozoa have been isolated and shown to contain intracellular volutin granules (3-7). Organisms as common as *E. coli* are capable of producing polyphosphate containing granules under proper environmental conditions (8). Srinath et al (9) concluded that protozoa of the *Epistylia sp.* could be related to excess phosphorus uptake in activated sludge systems. Feng (10) demonstrated that the operating conditions of the activated sludge process affected the rate and capacity of the process to remove phosphorus. Later, Sekikawa et al (11) studied the conditions which were responsible for the release of phosphorus by activated sludge. One of Sekikawa's conclusions was that lack of dissolved oxygen was a major cause of orthophosphate release. Levin (12) utilized the information obtained by these investigators, as well as the results of studies he performed (13), to produce a process capable of biochemically removing phosphorus from wastewater. This process has been named the PhoStrip^{R}* system.

This process utilizes the ability of microorganisms to take up and release phosphorus as a function of environmental conditions. Under aerobic conditions these organisms accumulate more phosphorus than that required for growth, while under an anoxic environment the excess phosphorus is released back to a liquid phase. Modifying the activated sludge process so that these characteristics could be used efficiently results in the concentration of phosphorus into a small sidestream, which can be economically treated with lime to precipitate phosphorus, as shown in Figure 1.

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* PhoStrip is a registered service mark of Union Carbide Corporation.

Early development of the process was undertaken by Biospherics, Inc. of Rockville, Md. This work resulted in a general outline of the operational parameters required for efficient phosphorus removal. Table 1 presents a summary of the type of phosphorus removal demonstrated by the process on a pilot scale. These data indicates that very efficient phosphorus removal may be achieved with chemical addition to the side stream flow.

This development work was highlighted by the full scale data obtained on a 12,100 m^3/d (3.3 mgd) plant at Seneca Falls, N.Y. (13) Conversion of the plant to the PhoStrip process was facilitated by the fact that, at the time of the study, the raw wastewater flow was only 3800 m^3/d (1 mgd). This enabled the plant to handle the full plant flow through one of the two reactor trains, thus freeing a primary clarifier to provide the anoxic environment required for the return sludge to release excess phosphorus. The tank providing this anoxic zone has been labelled the "stripper tank." The use of a single train to treat the entire wastewater flow also permitted the test to be run on a system that was much nearer the design hydraulic loading for the Seneca Falls plant. The use of the primary clarifier as the stripper tank was expeditious; however, it was significantly oversized relative to the wastewater flow entering the aeration basins. Table 2 presents a concise summary of the data obtained during the first thirty days of operation at Seneca Falls. This represents the period of operation during which the performance of the process was intensively monitored. The recycle flow rates presented are typical of a system utilizing sludge thickening in the stripper tank to produce the phosphate-enriched supernatant. The data on plant performance indicates that very high removals of both BOD_5 and phosphorus were achieved. The lime dosage is reflective of the chemical requirements anticipated for many full scale systems.

From testing of the process at Washington, D. C., it was determined that in the thickening mode of operation most of the phosphorus release occurred after the solid-liquid separation function was completed. Therefore, in order to achieve the supernatant phosphorus levels and the overall performance at Seneca Falls, the process was modified as shown in Figure 2. Sludge that had already released phosphorus was returned to the inlet line of the stripper, thus bringing the released phosphorus in contact with the liquid that would form the supernatant. This operational modification resulted in the ability to remove a significant fraction of the influent phosphorus through the supernatant stream leaving the top of the stripper tank. The remaining

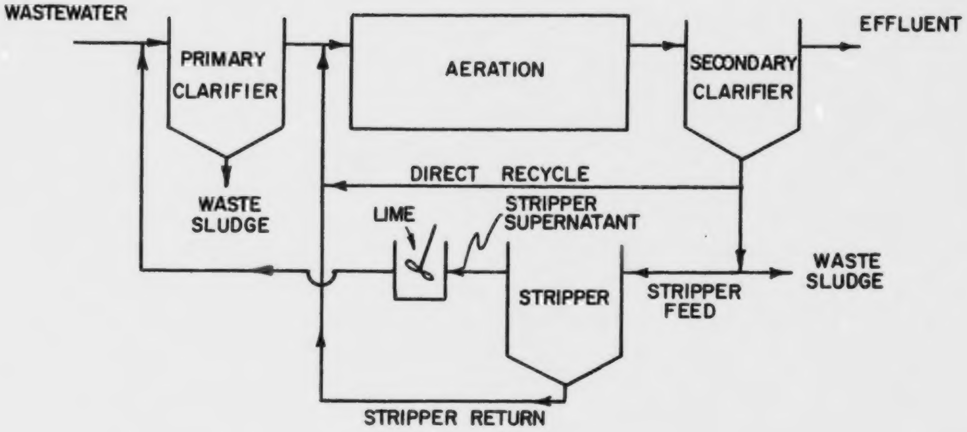


FIG. 1. Generalized PhoStrip schematic

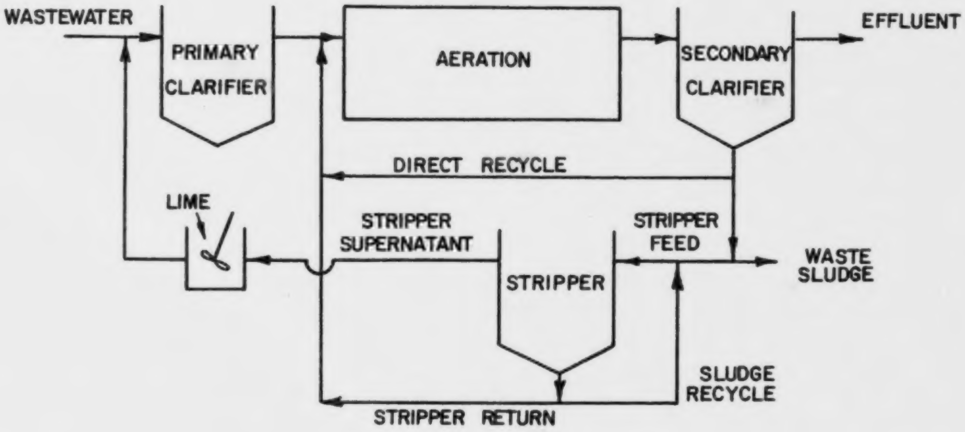


FIG. 2.. Sludge recycle option employed at Seneca Falls, N.Y.

TABLE 1. SUMMARY OF PHOSTRIP PILOT PLANT RESULTS

Results of Analysis of Raw Data Obtained from Biospherics, Inc.

Location	Phase Duration, days	Total Phosphorus		
		Raw Waste, mg/l	Effluent, mg/l	Removal, percent
Synthetic Waste	27	9.6	0.1	99
Washington, D. C.	30	6.8	0.8	88
Piscataway, Md.	30	5.1	0.5	90
Chicago, Ill.	10	3.0	0.3	90

TABLE 2. RESULTS OF PHOSPHORUS REMOVAL TEST AT SENECA FALLS, N. Y.

Plant Flow, m^3/d (mgd)	3400 (0.9)
Return Flows, percent of raw flow:	
Sludge to Stripper	24
Sludge to Aeration from Stripper	10
Supernatant to Primary Clarifier	14
Total Suspended Solids, mg/l:	
Mixed Liquor	1,440
Sludge to Stripper	7,840
Sludge to Aeration from Stripper	15,910
Influent, mg/l:	
BOD_5	158
Total Phosphorus	6.3
Effluent, mg/l:	
BOD_5	~ 4
Total Phosphorus	.6
Plant Performance:	
BOD_5 Removal, percent	98
Total Phosphorus Removal, percent	91
Lime Dose, Stripper Supernatant, mg/l of supernatant	170
Lime Dose, prorated to mg/l of raw flow	24

phosphorus removal was achieved with the waste biological sludge.

Early in 1974 Union Carbide Corporation began preliminary testing to determine the feasibility of this process. Initial studies were performed utilizing the flow scheme presented in Figure 1. A 0.95 l/s (15 gal/min.) pilot scale oxygen activated sludge system operating on Tonawanda, N.Y. wastewater served as the basic unit to which the PhoStrip process was added. This system was a four stage covered unit using pitch blade turbine surface aerators for mixing and mass transfer. High purity oxygen (99.5% O_2) was added to the first stage and the gas flowed cocurrently with the liquid from stage to stage. Both the mixed liquor and the oxygen depleted gas were discharged from the last stage of the reactor.

As with the work at Seneca Falls, the ability of the process to remove phosphorus depended on the ability to produce a supernatant stream with a high concentration of soluble phosphorus. It was found that low concentrations of dissolved oxygen or nitrate resulted in the inhibition of the release of phosphorus in the stripper tank. In general, the tests using the configuration shown in Figure 1 resulted in poor overall performance. Further testing utilizing the schematic presented in Figure 2 improved performance, but did not result in consistent high quality effluents. The nature of the system shown in Figure 2 was such that the supernatant from the stripper was produced through thickening. Because of solids flux limitations, this mode resulted in the operation of the secondary clarifier without a sludge blanket. The lack of a sludge blanket in the clarifier permitted significant amounts of oxygen to enter the stripper tank via the clarifier underflow. Since the sludge recycle scheme tends to make the stripper tank approach a completely mixed vessel, the dissolved oxygen coming in from the secondary clarifier may have been distributed well enough throughout the tank to significantly inhibit the rate of phosphorus release. However, as reported later on page 57, Peirano (31) successfully utilized this flow scheme at Reno/Sparks, Nevada.

To overcome the problems which resulted from a completely mixed stripper tank, the flow schemes presented in Figures 3 and 4 were tested. The modification shown in Figure 3 utilizes recirculation of stripper supernatant to wash or elutriate phosphorus from the solids leaving the stripper tank. Under this mode of operation, the sludge solids were allowed to progress through the tank in a plug flow manner while the supernatant and the liquid around the solids that was removed with the underflow solids tended to approach a completely mixed configuration. As with the sludge recycle system, the ability of this system to achieve equal concentrations of phosphorus in the supernatant and in the liquid surrounding the sludge leaving the stripper increased with increases in recirculation rates. At high recirculation rates a net upward flow of liquid occurs in the stripper tank, thus reducing the sludge thickness above the point of liquid injection. This coupled with the dissolved oxygen that entered the supernatant tended to reduce the efficiency of this technique although it was generally more efficient than the sludge recycle

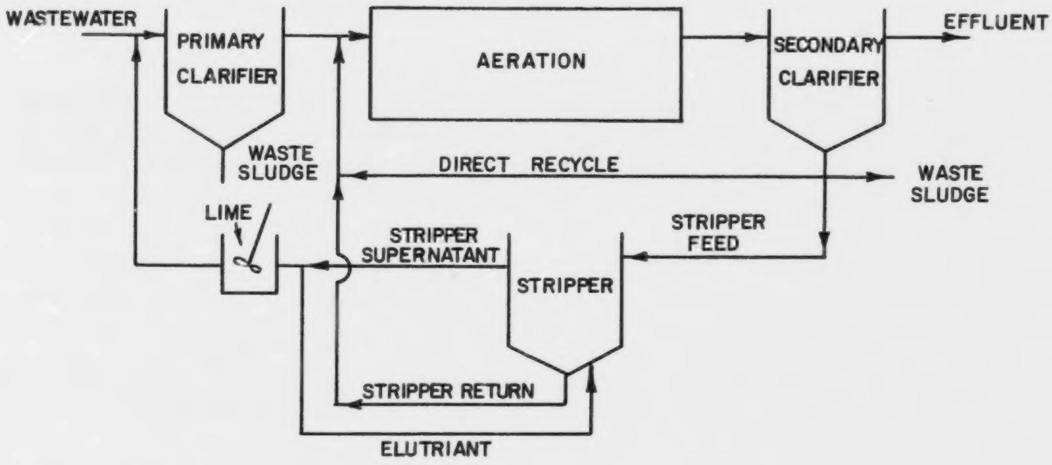


FIG. 3. Stripper supernatant elutriation option.

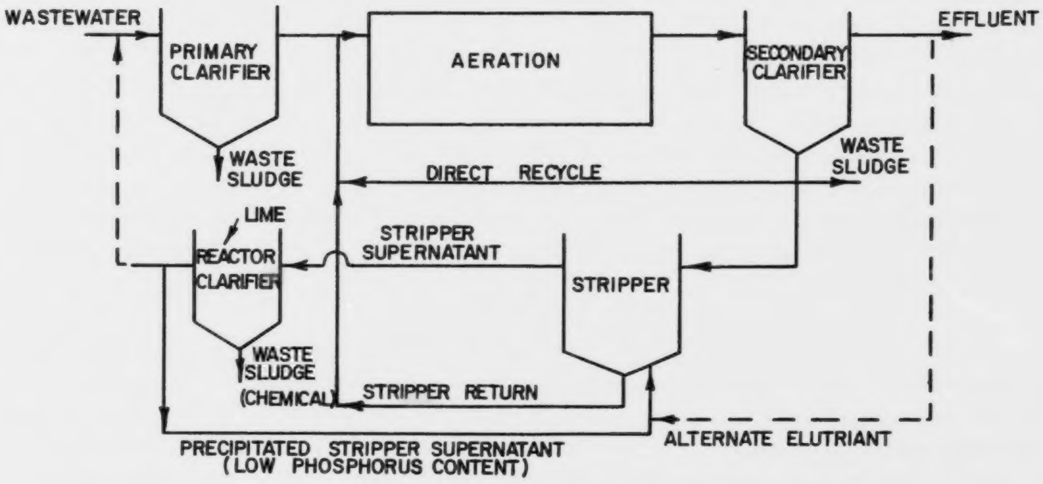


FIG. 4. Elutriation with liquids having low phosphorus concentrations. (precipitated stripper supernatant)

mode. The dissolved oxygen entered the elutriant through diffusional mass transfer at the air liquid interfaces at the surface of the tank and as a result of flow over the discharge weir of the tank. The first column of Table 3 summarizes typical data obtained utilizing the stripper supernatant recyle scheme.

Figure 4 presents a modified system as discussed by Matsch et al (14) and Drnevich et al (15). This system utilizes a liquid containing relatively low concentrations of both phosphorus and suspended solids to elutriate the sludge in the stripper and produce a supernatant high in dissolved phosphorus. In Figure 4 the elutriant is either the final effluent or the supernatant from the reactor clarifier which is used to perform the lime addition-precipitation function. This elutriation method overcomes the problems associated with the other techniques in that much lower flows are required to achieve equivalent elutriation efficiencies as defined in equation 1.

$$\varepsilon = \frac{\text{mass of phosphorus in supernatant}}{\text{total mass of phosphorus released in the stripper tank}} \qquad (1)$$

The performance of this system more closely approximates a countercurrent extraction system where plug flow characteristics are observed for both the liquid and solid phases in the stripper. In this configuration small concentrations of dissolved oxygen or nitrate in the elutriant have less of an effect on the rate of phosphorus release because only the solids in the bottom of the tank come in contact with the inhibitors. At that point, most of the phosphorus has already been released from the sludge and must merely be displaced to the supernatant. Table 3 presents a summary of two phases of operation where the system described by Matsch et al (14) was employed, compared to stripper supernatant recirculation. During the phase represented by the data in column two, the elutriant employed was supernatant from the stripper after the phosphorus was removed via lime addition. The last column presents data from a system employing final effluent as the elutriant. In comparing the performance of these systems with the data from the process employing stripper supernatant recirculation, it is important to note that countercurrent type systems produce higher elutriation efficiencies at significantly lower elutriation flow rates.

Similar testing of the process on air activated sludge pilot plants on synthetic and Tonawanda wastewaters confirmed the superiority of the countercurrent elutriation type system with respect to the ability to achieve high stripper supernatant and low final effluent phosphorus concentrations. This system also eliminates the need for thickening in the stripper tank because the supernatant is formed from the elutriant. This reduced dependency on thickening permits the stripper tank to be designed with significantly lower cross sectional area and volume.

Subsequent to the initial testing of the process at Tonawanda, N.Y.,

TABLE 3. PILOT PLANT OXYGEN ACTIVATED SLUDGE SYSTEM AT TONAWANDA, N. Y.

Mode of Operation	Stripper Supernatant Used for Elutriation	Precipitated Stripper Supernatant Used for Elutriation	Final Effluent Used for Elutriation
Duration of Phase, days	12	20	23
Feed Rate, l/s (gpm)	.76 (12)	.95 (15)	.95 (15)
Recycle Rate, % of Q	25.	40.	33.
Stripper Feed, % of Q	25.	19.	20.
Stripper Return to Aeration, % of Q	12.	18.	19.
Stripper Supernatant Rate, % of Q	13.	9.	11.
Elutriation Rate, % of Q	25.	8.	9.
Anaerobic Retention Time, hr. (based on stripper return to aeration)	9.9	6.1	6.3
Total Suspended Solids, mg/l:			
Mixed Liquor	2,929	5,189	4,400
Stripper Feed	14,740	18,600	17,500
Stripper Return	31,590	18,140	17,540
Effluent	10	7	10
BOD_5, mg/l:			
Influent	78	100	120
Effluent	10	10	18
Total Phosphorus, mg/l:			
Influent	5.3	7.5	11.1
Effluent	0.5	0.5	0.8
Stripper Return (soluble)	59	36	60
Stripper Supernatant	35	58	78
Elutriation Efficiency (ε)	.40	.45	.43

the performance capabilities of the process have been demonstrated via pilot plant programs on wastewaters from Texas City, Texas; Adrian, Michigan; Southtowns, N.Y.; Brockton, Mass. (16); and Findlay, Ohio. Table 4 contains some of the important characteristics and data obtained from the various studies. Each of the programs employed a variation of the countercurrent elutriation system. The Texas City full scale plant has a State-imposed effluent phosphorus limitation of less than 2.0 mg P/l, and the BOD_5 and suspended solids levels have to be maintained at less than 10 mg/l. The data in Table 4 shows that the phosphorus removal was achieved with no difficulty, but the effluent would have to be filtered because the effluent suspended solids achievable by activated sludge systems are higher than the State's limitation.

As is the case at Texas City, the activated sludge system in operation at Adrian is not able to reduce effluent suspended solids to meet State standards. Therefore, tertiary filtration will be employed. Since 88% of the effluent phosphorus is in the form of suspended solids, tertiary filtration to less than 10 mg/l suspended solids would result in an effluent total phosphorus of less than 1 mg/l.

The pilot plant program performed on the wastewater from Southtowns, N.Y. differs from the previously discussed studies in that the pilot plant was run at Tonawanda, N.Y. as opposed to an on-site study. This required periodic trucking of wastewater from Southtowns to Tonawanda. The raw feed was stored for up to three days at ambient temperatures before being fed to the reactor. This resulted in extremely septic conditions which may have been in part responsible for the effluent suspended solids problems. As with the other locations, final effluent filtration will be required to achieve a 10/10 (BOD_5/suspended solids) standard. At an effluent suspended solids level of 10 mg/l an effluent total phosphorus level of less than 0.5 mg P/l is anticipated for the full scale Southtowns effluent.

The study at Brockton, Mass. utilized larger equipment than the other studies and, as a result, was less subject to sampling and operational problems. The Brockton data shows that, at effluent suspended solids levels typical of activated sludge systems, effluent phosphorus levels below 1 mg P/l are achievable without filtration. The low temperature data of the Brockton study and the high temperature data of the Adrian study indicated that the process could perform well over a range of temperatures.

The Findlay, Ohio study represents one of the first attempts to run the process with the contact stabilization or step feed modification of the activated sludge process. After a large amount of background data was obtained to quantify the interactions between this type of activated sludge process and the PhoStrip processes, the operating data presented in Table 4 were obtained.

TABLE 4. RESULTS OF PILOT SCALE DEMONSTRATION PROGRAMS

	Location				
	Texas City, TX	Adrian, MI	Southtowns, N.Y.	Brockton, MA	Findlay, OH
Type of Activated Sludge Pilot Unit	conventional air	conventional air	conventional air	conventional air	step feed
Type of Elutriant	primary effluent	primary effluent	primary effluent	primary stripper supernatant	final effluent
Type of Influent	municipal	municipal	municipal	municipal	municipal
Duration of Phase, days	14	31	19	26	12
Feed Rate (Q), l/s	.011	.009	.0083	.24	.0073
Recycle Rate, % of Q	14	20	17	37	30
Stripper Feed Rate, % of Q	14	17	7	20	22
Stripper Return to Aeration, % of Q	12	15	7	18	20
Stripper Supernatant, % of Q	17	17	1.3	14	20
Elutriation Rate, % of Q	15	15	1.3	12	20
Influent Characteristics, mg/l:					
BOD_5 (COD)	73	61	121	212	(315)
Total Phosphorus	7.9	8.4	3.0	11.3	6.9
Total Suspended Solids	69	74	89	344	62
Temperature (Aeration), °C	22.7	26.7	18.9	12.2	13.0
Total Suspended Solids (Mixed Liquor), mg/l	1,672	2,244	2,980	2,525	2,000
F/M (BOD_5/day/kg MLVSS)	0.4	0.3	0.4	0.5	0.6 (COD basis)
Effluent Characteristics, mg/l:					
BOD_5 (COD)	29	16	34	11	(45)
Total Effluent Suspended Solids	25	35	51	15	11
Total Phosphorus	0.7	1.6	1.4	0.8	0.7
Ortho Phosphorus	0.3	0.2	<0.1	0.3	0.1

OBJECTIVES OF PROGRAM

The purpose of the program at Reno/Sparks, Nevada was to demonstrate the performance of the PhoStrip process on a full scale system and establish overall process design relationships for the system. Pilot scale information was to be used to establish the performance characteristics of a system designed specifically to incorporate the process as opposed to a system utilizing existing unmodified tankage. A comparison between the pilot and full scale system was designed to establish the importance of designing the activated sludge and PhoStrip systems to complement each other. Information on lime requirements, sludge production rates, sludge dewatering characteristics and aerobic and anaerobic sludge digestion was also to be determined.

SCOPE OF PROGRAM

The original scope of the project was to test the process on 50,100 m^3/d (13 mgd) and 76,000 m^3/d (20 mgd). These data were to be compared to the results previously obtained at 25,000 m^3/d (6 mgd) and information generated via pilot scale testing. However, the duration of the study was shortened due to an unforeseen U.S. EPA Region IX one-year-long water quality study on the Truckee River, which is the discharge point for the effluent from the Reno/Sparks treatment plant. The study required that all inputs to the river be characteristic of standard operation, thus precluding any test work.

Before the water quality study started, a two-month operational period at winter operating conditions generated enough data to evaluate the 50,100 m^3/d (13 mgd) case and establish significant relationships for the design and control of the process. The 76,000 m^3/d (20 mgd) case was never tested. Chemical sludge production, lime requirements, and sludge dewatering characteristics were obtained via pilot scale operation.

The proposed studies on the aerobic and anaerobic digestion of the combined sludges were not completed due to the termination of the test program.

SECTION 2

CONCLUSIONS

The PhoStrip system has demonstrated the capability of reducing effluent total phosphorus concentrations to less than 1 mg/l on both the pilot and full scale systems tested at Reno/Sparks, Nevada.

The phosphorus removal was achieved through the ability of activated sludge to take up and release phosphorus as a function of being cycled through aerobic and anoxic zones. The activated sludge from the PhoStrip system generally contains a higher phosphorus content than sludge from conventional activated sludge systems and, thus, is capable of removing more phosphorus with the waste biomass. The anoxic tank produces conditions which cause the activated sludge to release phosphorus into a small side stream. The phosphorus from this side stream can be economically precipitated with lime.

To achieve low effluent phosphorus levels, it is necessary to achieve both a low filtered effluent phosphorus level and a low particulate effluent phosphorus content. The effect of particulate matter on effluent phosphorus concentrations can be controlled by controlling effluent suspended solids levels and/or phosphorus concentration in the suspended solids.

The control of effluent suspended solids levels below that normally obtained by activated sludge systems can be achieved through filtration. The control of the phosphorus content in the suspended solids and the control of the effluent filtered phosphorus concentration is highly dependent on the amount of phosphorus that is released in the stripper and the ability to separate the released phosphorus from the sludge in the stripper tank. The amount of phosphorus released depends on the operating conditions in the stripper relative to those in the aeration basin. The ability to separate phosphorus from the sludge depends on the method employed. Both the sludge recycle and the elutriation systems utilized in this program proved to be viable methods of achieving the required phosphorus/sludge separation. The elutriation method gives the designer more flexibility and the ability to design the process with minimal stripper tank volume requirements.

The amount of lime required for the process depends on the quantity of stripper supernatant (usually between 5 and 15% of the influent flow) and the pH of precipitation. A pH of approximately 9.0 is sufficient to precipitate most of the phosphorus from the supernatant. Utilizing this pH results in the production of relatively small quantities of sludge which is high in phosphorus content. This chemical sludge appears to be ideally suited as a high phosphate fertilizer.

Economic evaluations of the process indicate that it is a highly cost effective alternative to phosphorus removal by mineral addition.

SECTION 3

RECOMMENDATIONS

A by-product of the PhoStrip system is a high phosphorus containing calcium carbonate type sludge which has potential value as a fertilizer. Studies should be undertaken to determine the characteristics of this sludge with respect to its applicability to farm land.

The PhoStrip system represents one aspect of a multifaceted sewage treatment plant. The incorporation of this process into the overall wastewater treatment plant should be explored utilizing a systems approach.

SECTION 4

PROCESS THEORY

BIOCHEMICAL MECHANISMS

The performance of the PhoStrip process is based on the formation and elimination of intracellular polyphosphate granules (volutin). This phenomenon of polyphosphate formation and decay is induced through the continuous cycling of activated sludge through aerobic and anoxic conditions. As a consequence of the aerobic environment, the conditioned microorganisms remove phosphate from solution and produce the storage product, volutin. These organisms then release phosphate into solution while under a condition of stress (anoxic surroundings). The precise biochemical pathways of this phenomenon have not yet been delineated. However, the hypotheses that follow have been developed based on work performed on bacteria isolations (17) and the response of the process to stimuli (temperature, substrates, etc.).

Fuhs and Chen (17) isolated organisms capable of excessive phosphorus uptake from PhoStrip systems operating at Baltimore, Md. and Seneca Falls, N.Y. Of these isolates, the organisms of the genus Acinetobacter most vigorously demonstrated rapid uptake of phosphorus under aerobic conditions and rapid reduction in the size of polyphosphate granules while anoxic. Russ (18) isolated an organism which he named Acinetobacter phosphadevorus from the Rilling Road Sewage Treatment Plant, San Antonio, Texas. This organism showed the same phosphorus storage potential as those isolated from the PhoStrip systems. It is significant to note that these organisms are obligate aerobes which effectively compete with the facultative species in a system with prolonged anoxic periods. It is hypothesized that the formation of the volutin granules during the aerobic portion of the PhoStrip process is largely responsible for this ability to compete.

Investigations into the purpose of polyphosphate formation have indicated the following functions for this storage product. (19) First, polyphosphate may be used as a phosphate reserve. This hypothesis would suggest that the major purpose of polyphosphates is to supply phosphorus for metabolism during periods of phosphate starvation. In situations where phosphorus is stored as a result of undergoing periods of phosphorus starvation, this may be the case. However, there is no period during which the organisms are starved of soluble phosphorus in the PhoStrip process and, therefore, no need to store phosphorus according to the rationale of this hypothesis. The second major speculation about the function of polyphosphate contends that polyphosphate is important for the regulation of the phosphorus economy of the cell. It is important to maintain the inorganic phosphate (Pi) level of the cell below a certain level to insure efficient energy usage and permit other metabolic reactions to occur. This hypothesis may indeed explain why polyphosphate is stored under aerobic conditions but it yields no clue as to why it is released during anoxic periods. To explain phosphorus release as a result of the "Pasteur effect" is inconsistent with the observation that obligate aerobes (genus Acinetobacter) release phosphorus without undergoing fermentation reactions. The explanation that phosphatase enzymes may be responsible for polyphosphate reduction may eventually satisfy the shortcomings of this hypothesis. However, no linkage has yet been developed to explain why the presence of facultative bacteria is necessary for the obligate aerobes to release significant quantities of phosphorus. (17)

The third suggested use of polyphosphate is as an energy source, either as a "phosphagen" (20-21) or through direct substitution for adenosine triphosphate (ATP) (22) in energy requiring reactions. A broad definition of a phosphagen is a naturally-occurring compound which stores phosphate bond energy and from which phosphoryl groups can be transferred to adenosine diphosphate (ADP) to produce ATP.

In an anoxic environment the energy source hypothesis would predict that obligate aerobes would utilize energy stored in polyphosphates to continue some of the metabolic functions of the cell. Equation (2) presents an example of a possible use of energy to produce the Krebs Cycle precursor, acetyl Coenzyme A, from acetate. The acetate necessary to drive this

$$\text{acetate} + \text{ATP} \longrightarrow \text{acetyl Coenzyme A} + \text{ADP} + \text{Pi} \qquad (2)$$

reaction to completion is produced by the facultative bacteria while under the anoxic conditions of the stripper tank. Experiments performed at Union Carbide's Tonawanda labs (23) and Penn State University where acetates were added to sludges obtained from PhoStrip systems have indicated enhanced phosphorus release while the addition of glucose was significantly less effective at increasing the rate of phosphorus release from the cells. The importance of acetate formation is enhanced by the fact that the organisms of the genus Acinetobacter prefer acetates as a substrate and will not metabolize glucose. Since the Krebs cycle cannot function without either molecular oxygen or nitrate as the final electron acceptor, the acetyl Coenzyme A content of the cell increases while the ratio of ATP to ADP decreases. ATP may also be used as the energy source for the formation of other carbonaceous storage products. The low ratio of ATP to ADP will cause the polyphosphokinase enzyme (20) to utilize the bond energy and the phosphoryl group **stored in** the polyphosphate to produce ATP from ADP. Equation (3) is reversible, and the net direction of the reaction is dependent on the relative concentrations of ATP and ADP. This reaction is slow relative to the ATP utilizing reactions; therefore, no ATP buildup is possible during the anoxic period. Alternately, when the ATP level of the

$$ADP + PP_n \rightleftarrows ATP + PP_{n-1} \tag{3}$$

where PP_n = polyphosphate

cell is diminished as in reaction (2), it is possible that short chain polyphosphates replace ATP as shown in equation (4). (21) Either of

$$\text{acetate} + PP_n \rightleftarrows \text{acetyl Coenzyme A} + PP_{n-1} + Pi \tag{4}$$

these reactions may be predominantly responsible for the polyphosphate reduction observed (17,24) in PhoStrip sludges held under anoxic conditions. Investigators at Penn State University have also established a significant reduction in ATP levels in sludges leaving the anoxic PhoStrip stripper tank. (24)

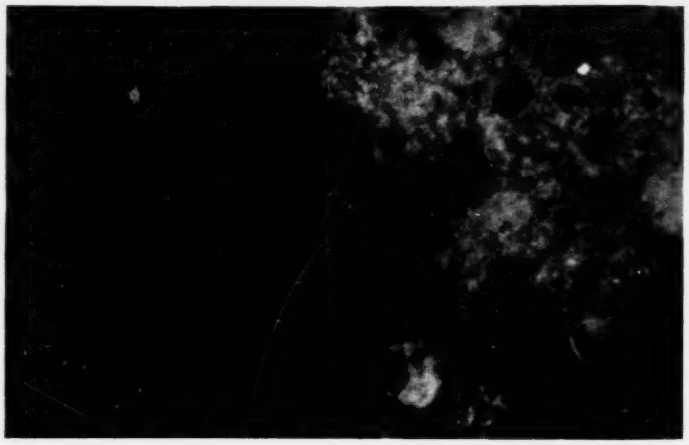
According to this hypothesis, the sludge leaving the stripper tank may be characterized by the following conditions:

1. low ATP/ADP ratio,
2. high concentrations of acetyl Coenzyme A and/or other carbonaceous storage products,
3. relatively low polyphosphate levels in the sludge.

The effect of the anoxic period on the polyphosphate level in the sludge is shown in Figure 5. These photomicrographs are sludge samples which were stained to highlight the volutin granules within the bacteria. The results of this test show that the sludge entering the stripper contained many large clusters of polyphosphate containing organisms while the same sludge shows a significant reduction in the size and quantity of the polyphosphate granules after only an 8-hour anoxic period.

Upon return to the aerobic basin of the PhoStrip system, the organisms rapidly produce ATP because the reaction rates in the Krebs cycle are enhanced by the low ATP/ADP ratio and the high concentration of the storage products. After a short period of time, these reactions increase the ATP/ADP ratio, thus driving the reaction in equation (3) from right to left. According to this

Before entering the stripper



After leaving the stripper

Figure 5. Sludge polyphosphate granules before and after anoxic period in stripper tank (dark areas are clusters of polyphosphate containing bacteria.

mechanism, the uptake rate of phosphorus in the aeration basin should be zero order with respect to phosphorus concentration in solution. Experiments performed at Union Carbide indicate zero order uptake rate at phosphorus levels greater than 0.5 mg ortho phosphorus/liter of solution. (25)

Consistent with the above hypothesis, early testing performed by Biospherics, Inc. (26) and later studies (27) have indicated that phosphate release is inhibited in the stripper tank by low levels of nitrate and dissolved oxygen. The presence of nitrate and oxygen stimulates the ATP producing reactions of the Krebs cycle.

The major argument against the phosphagen mechanism results from work by Harold (28) on fungi. He was not able to see polyphosphate breakdown when energy generation was limited or ATP generation from the polyphosphate. However, the system that he studied was not consistent with that responsible for phosphorus removal in the PhoStrip system. Polyphosphate granules produced via the over-plus phenomenon (phosphate starvation) and growth-inhibiting situations (acidity, sulfur starvation) may serve different functions than the polyphosphate produced by the organisms in the PhoStrip process.

PHOSPHORUS MASS BALANCE CONSIDERATIONS

Figure 6 presents a phosphorus mass balance around a theoretical PhoStrip system. As shown in this figure, there are three outlet locations for the influent phosphorus: the final effluent, the waste activated sludge, and the waste chemical sludge. Ideally, all the removal of phosphorus would be with the waste activated sludge since this would result in no chemical requirements for the process. In certain instances this has been achieved with the PhoStrip process and can be used for the design of some full scale systems. However, with the amount of sludge wasting expected in the theoretical case of Figure 6, 342 kg/day of phosphorus would have to be present with the volatile suspended solids of the waste activated sludge. This represents more than 16% by weight of the waste volatile suspended solids. The maximum observed in PhoStrip systems has been only 8%. Even if 8% phosphorus levels were utilized for design purposes, the effluent quality of the process would be limited by the effluent volatile suspended solids (EVSS) concentration. At 0.08 P/VSS, the phosphorus in the effluent due to 23 mg EVSS/l would be about 1.8 mg/l. To achieve an effluent phosphorus level of less than 1.0 mg/l, EVSS levels near 10 mg/l would be required. In most cases, effluent filtration would be necessary to maintain such low suspended solids levels. The goal of complete phosphorus removal via activated sludge wasting is more readily achievable on systems with higher sludge wasting rates and/or lower influent total phosphorus levels than shown in Figure 6.

The use of lime to insolubilize phosphorus permits the PhoStrip process

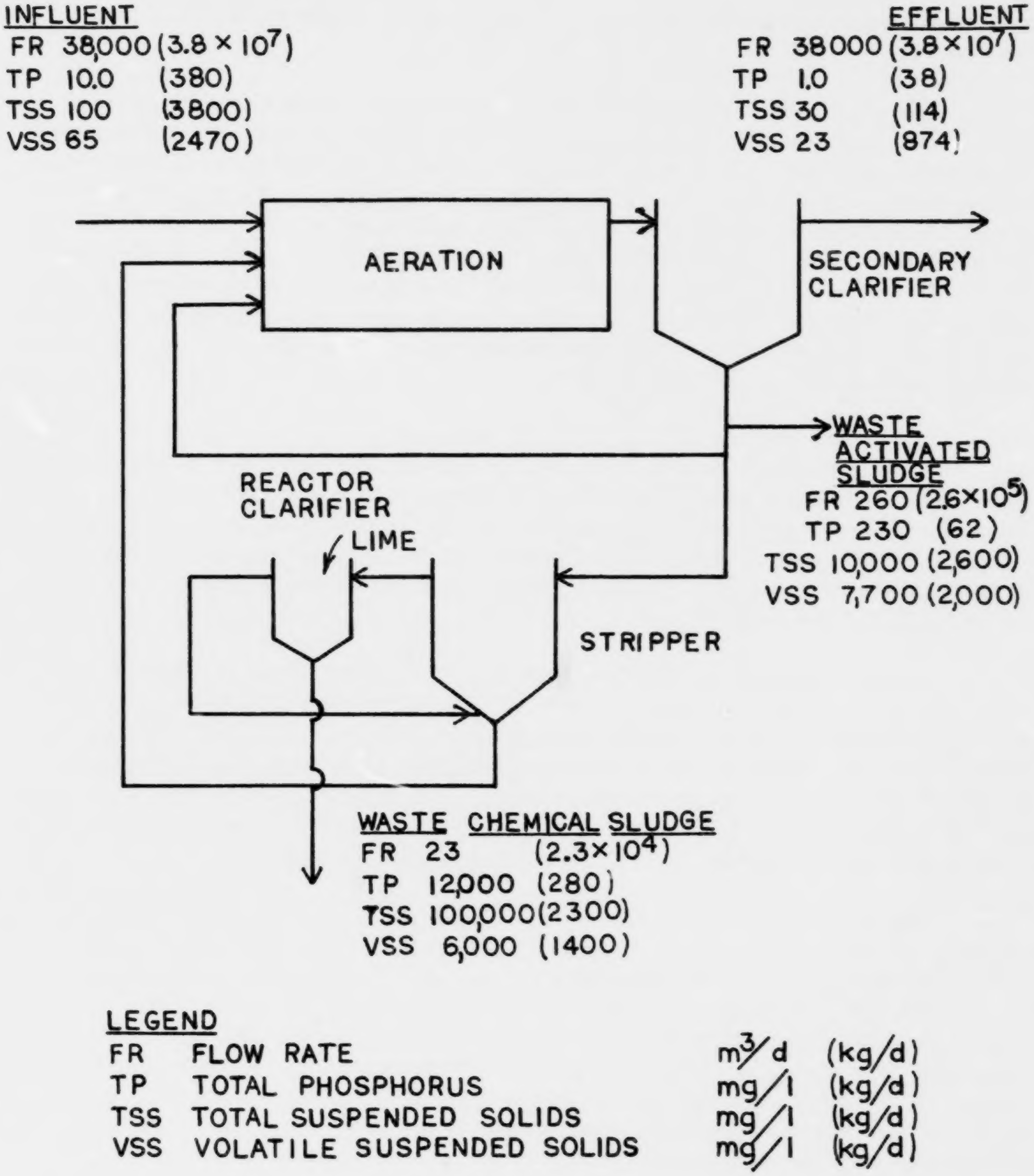


Figure 6. Phosphorus mass balance - 38,000 m^3d (10 mgd) plant size.

to be designed at any P/VSS ratio between 0.03 and 0.08 in the sludge from the secondary clarifier. At the P/VSS ratio employed in Figure 6 (0.03), effluent phosphorus levels of less than 1.0 mg/l can be achieved without post filtration.

The phosphorus balance presented in Figure 7 represents conditions expected in streams entering and leaving the stripper tank for the theoretical situation developed in Figure 6. Evaluation of this information indicates that only a 0.007 differential in the P/VSS ratio of the sludge entering and leaving the stripper is required to achieve efficient performance. The P/VSS level in the stream entering the stripper drops from 0.030 to 0.023 in the stream returned to aeration. A given level of P/VSS reduction requires a minimum mass flow of VSS through the stripper so that mass balances are satisfied. Utilizing a small differential in the P/VSS levels in the sludge permits the use of short anoxic periods in the stripper tank.

Since phosphorus is continuously released while the sludge is in the stripper tank, a portion of this solubilized material is recycled to the aeration basin with the return flow. Elutriation efficiency is a measure of the fraction of solubilized material that is available for precipitation as well as the fraction that is returned to the aeration basin. At a 45% elutriation efficiency, more than half of the solubilized phosphorus returns to the aeration basin to be taken up by the sludge.

Two methods of maximizing the amount of phosphorus in the supernatant have been presented earlier. The sludge recycle method (Figure 2) and low phosphate elutriation modification (Figure 4) were discussed briefly. The sludge recycle system tends to make the stripper tank approach a completely mixed basin when the recycle rate is high. Figure 8 shows the flow schematic for a stripper tank in the sludge recycle mode as well as the assumptions for a theoretical mass balance analysis of this mode of operation.

Figure 9 summarizes the results of the mass balance analysis of the sludge recycle option. High elutriation efficiencies are only achieved at high recycle fractions and high supernatant flow rates. The results of this analysis are consistent with pilot scale data obtained on the PhoStrip process. The elutriation efficiency for the low phosphate containing elutriant was also obtained on a pilot scale. The efficiency of this system is also presented in Figure 9. At relatively low elutriation rates, the low phosphate elutriation (LPE) system approaches the performance of the sludge recycle option. As the elutriation rate increases, the performance of the LPE system improves substantially over that of the sludge recycle system. In properly designed systems, the elutriation rate is equal to the stripper supernatant flow. The effects of these relationships will be established in an example which will be presented later.

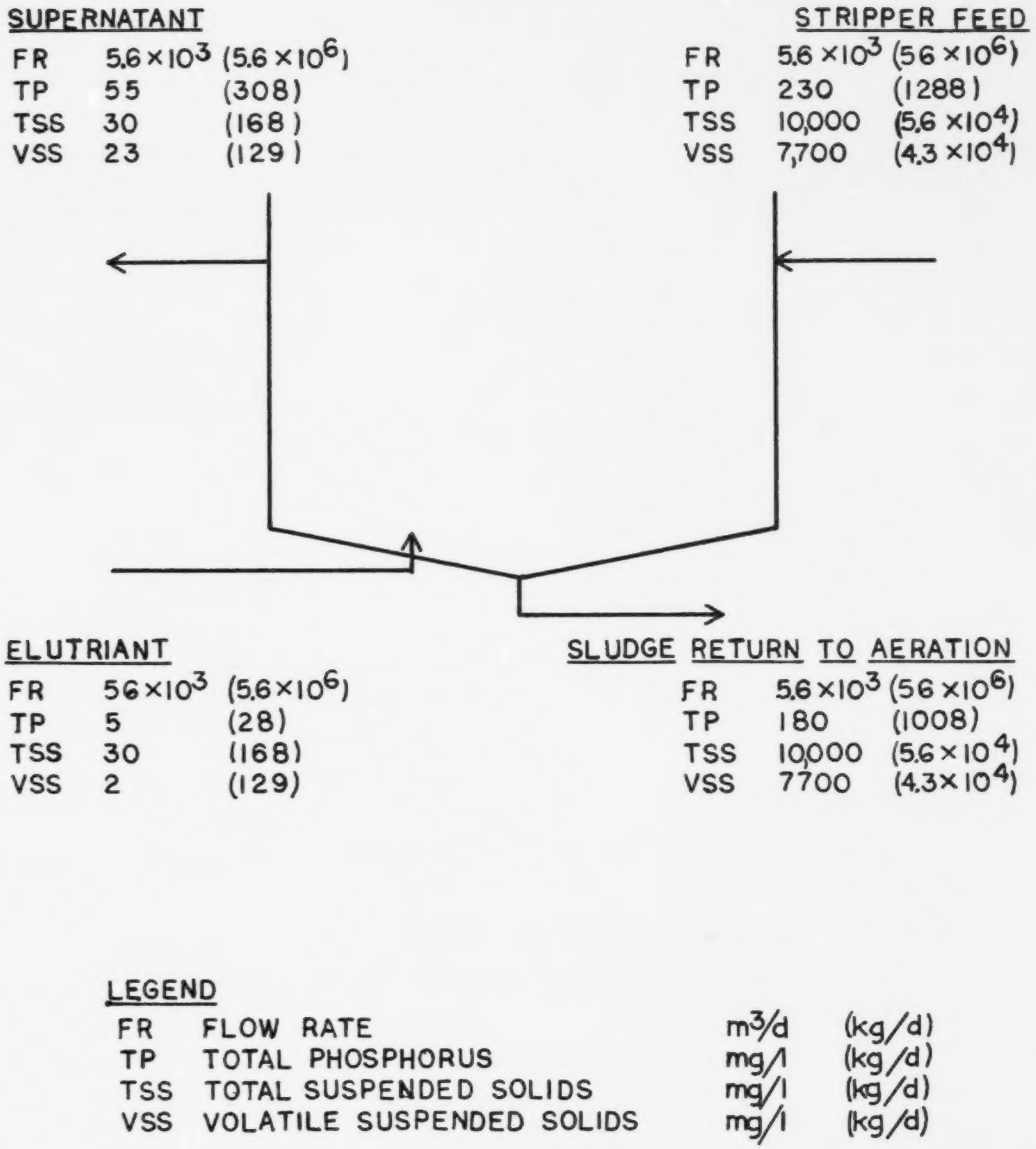
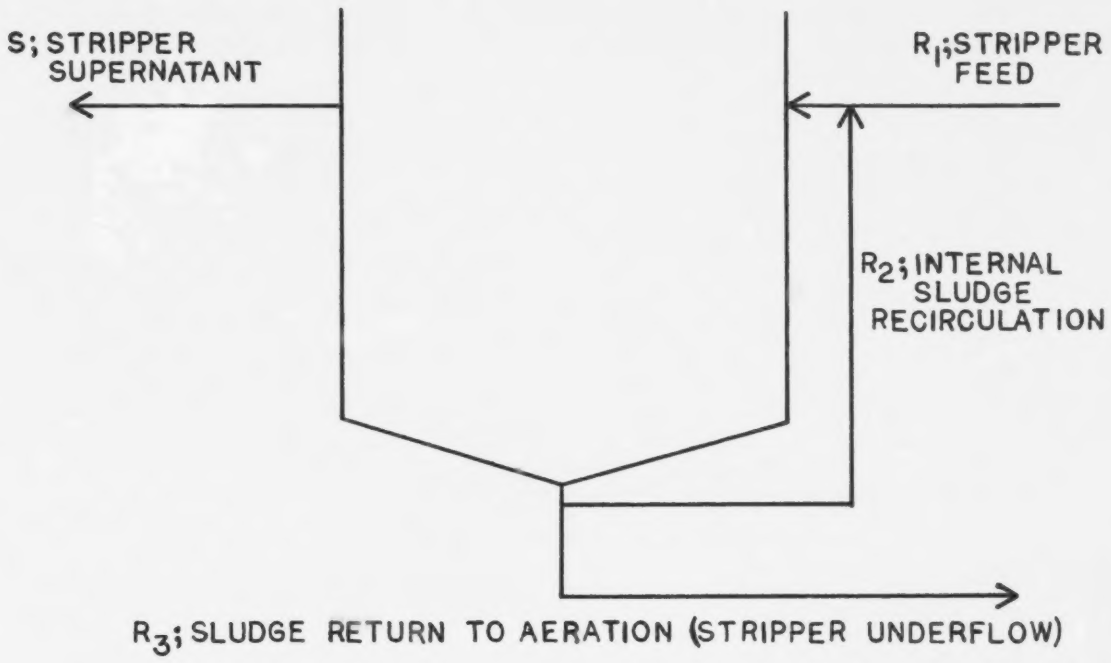

SUPERNATANT
FR 5.6×10^3 (5.6×10^6)
TP 55 (308)
TSS 30 (168)
VSS 23 (129)
STRIPPER FEED
FR 5.6×10^3 (56×10^6)
TP 230 (1288)
TSS 10,000 (5.6×10^4)
VSS 7,700 (4.3×10^4)
ELUTRIANT
FR 56×10^3 (5.6×10^6)
TP 5 (28)
TSS 30 (168)
VSS 2 (129)
SLUDGE RETURN TO AERATION
FR 5.6×10^3 (56×10^6)
TP 180 (1008)
TSS 10,000 (5.6×10^4)
VSS 7700 (4.3×10^4)
LEGEND
FR FLOW RATE m^3/d (kg/d)
TP TOTAL PHOSPHORUS mg/l (kg/d)
TSS TOTAL SUSPENDED SOLIDS mg/l (kg/d)
VSS VOLATILE SUSPENDED SOLIDS mg/l (kg/d)

Figure 7. Mass balance around the stripper tank.


S; STRIPPER SUPERNATANT
R_1; STRIPPER FEED
R_2; INTERNAL SLUDGE RECIRCULATION
R_3; SLUDGE RETURN TO AERATION (STRIPPER UNDERFLOW)

ASSUMPTIONS:

1. NO INHIBITION OF PHOSPHORUS RELEASE DUE TO PRESENCE OF NITRATES OR DISSOLVED OXYGEN IN R_1.
2. R_1 AND R_2 ARE COMPLETELY MIXED.
3. THE SOLUBLE PHOSPHORUS CONCENTRATION IN R_1 IS NEGLIGIBLE.

Figure 8. Stripper tank mass balance, sludge recycle modification.

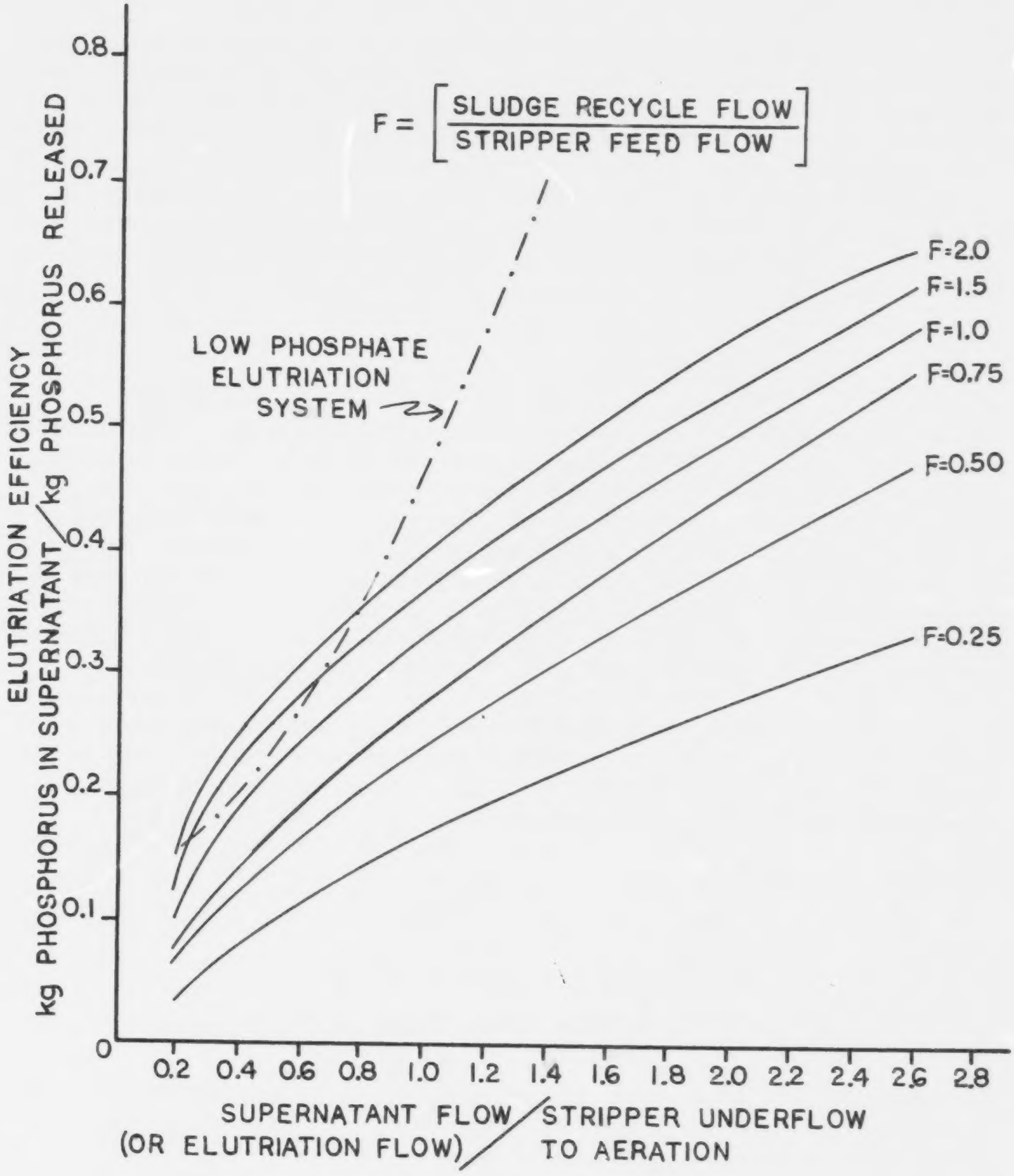


Figure 9. Elutriation efficiency for the sludge recycle and elutriation systems.

SOLIDS FLUX THEORY IN RELATION TO STRIPPER DESIGN

The theoretical evaluation of final settling tank behavior with respect to thickening has been performed by Dick. (29) Generally, the total flux of suspended solids in a settling tank or thickener (G_T) can be summarized as the sum of the flux due to gravity settling (G_g) plus that due to the downward bulk velocity as a consequence of the withdrawal of solids from the bottom of the clarifier (G_u). Equation (5) presents a mathematical description of the total flux as a function of suspended solids concentration at the point of interest, c_i, the settling velocity at that concentration, v_i, and the bulk velocity, u.

$$G_T = c_i v_i + c_i u \qquad (5)$$

Figure 10 graphically illustrates the addition of the two types of flux curves to produce the total flux. With reference to Figure 10, if the suspended solids concentration of a stream entering the settling tank, c_o, is greater than c_m, and the desired underflow concentration is c_u, then the sludge in the settling tank may exist at all concentrations between c_m and c_u. However, one concentration, c_L, exhibits the minimum solids flux called the limiting flux, G_L. For a given surface area G_L depends only on the settling characteristics of the sludge and the underflow rate. For proper steady state thickening performance, the solids loading per unit area, G_A, must equal the limiting flux.

In designing a stripper tank for the PhoStrip process, the above requirement provides the means for determining the cross-sectional area necessary to obtain a given underflow concentration, c_u. Referring to Figure 8, equation (6) predicts the area required to give the desired underflow

$$A = \frac{C_1 R_1 + C_2 R_2}{G_L} \qquad (6)$$

C_1 = suspended solids concentration in stream R_1

C_2 = suspended solids concentration in stream R_2

concentration for the sludge recycle system. As was presented earlier, there is a minimum mass of sludge, $C_1 R_1$, which must enter the stripper tank to satisfy mass balance requirements for phosphorus release. Mass balances also dictate a relationship between supernatant volume and recirculation rates (Figure 9) necessary to achieve high elutriation efficiency. The limiting flux in equation (6) is defined in equation (7). Therefore, once R_1 is defined

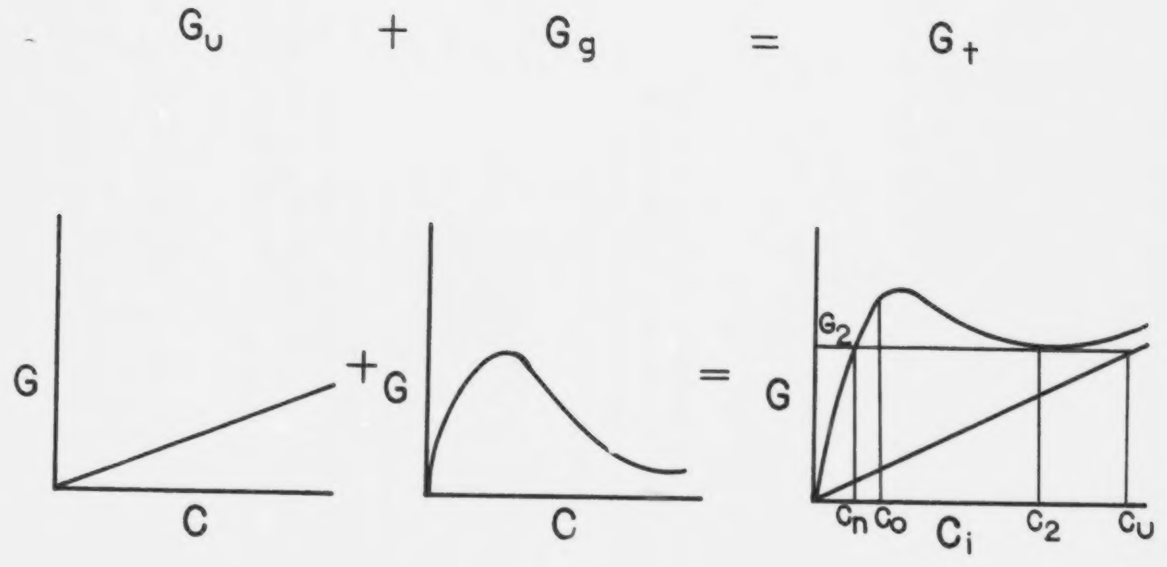


Figure 10. Components of total flux settling tanks.

$$G_L = C_L v_L + C_L \frac{R_2 + R_3}{A} \qquad (7)$$

C_L = limiting flux suspended solids concentration

v_L = limiting flux settling velocity

from the suspended solids level in the sludge from the secondary clarifier and mass balance considerations, the cross-sectional area of the stripper depends only on the settling characteristics of the sludge and the tradeoff made between the stripper supernatant flow and the recirculation rate according to Figure 9. Equation (8) demonstrates this dependence on the recirculation rate and settling characteristics (v_L). It should be noted that $C_2 = C_u$ and since C_L is always

$$A = \frac{C_1R_1 + C_2R_2 - C_L(R_1 - S + R_2)}{C_Lv_L} \qquad (8)$$

less than C_u, the cross-sectional area is minimized by reducing R_2. However, with the sludge recycle system, R_2 can only change if S is changed. A balance between the cost of lime for precipitation resulting from changes in S and the capital cost associated with cross-sectional area must be struck.

For the sake of consistency, it is important that a solids flux analysis of the secondary clarifier be performed for the proposed operating conditions. This is necessary because one of the key parameters for designing the area and volume of the stripper tank is the suspended solids level in the recycle stream from the secondary clarifier (C_1). Dick and Young (30) presented a mathematical method of determining the performance of secondary clarifiers from a thickening viewpoint. They utilized the relationship presented in equation (9) to quantify the effect of suspended solids concentration on settling velocity, v_i. The constants, a and n, were shown to depend on the physical

$$v_i = a\,C_i^{-n} \qquad (9)$$

characteristics of the sludge. The effective prediction of settling velocity via equation (9) is limited to a range of solids concentrations bounded by approximately 3,000 mg/l and 30,000 mg/l. This range may vary, depending on the physical characteristics of the sludge. Dick and Young (29) showed that utilizing equation (9), the limiting flux could be determined through the application of equation (10). For the PhoStrip process

$$G_L = \left[a(n-1)\right]^{1/n}\left[\frac{n}{n-1}\right]\left[\frac{rQ}{A}\right]^{\frac{n-1}{n}} \tag{10}$$

r = the sludge recycle flow divided by the raw feed flow

Q = the raw feed flow

$\frac{rQ}{A}$ = u

$u = \frac{R_2 + R_3}{A}$, thus equation (11) results from substitution.

$$G_L = \left[a(n-1)\right]^{1/n}\left[\frac{n}{n-1}\right]\left[\frac{R_2 + R_3}{A}\right]^{\frac{n-1}{n}} \tag{11}$$

G_L can also be defined according to equation (12).

$$G_L = \left(\frac{R_2 + R_3}{A}\right) C_3 \tag{12}$$

Combining equations (11) and (12) and solving for A, yields equation (13).

$$A = \frac{R_2 + R_3}{\left\{\frac{\left[a(n-1)\right]^{1/n}\left[\frac{n}{n-1}\right]}{C_3}\right\}^{n}} \tag{13}$$

where $C_3 = C_u$

Equation (13) is used for determining the cross-sectional area required for a given set of a and n constants related to the sludge in question. The R_2, R_3, and C_3 variables are determined from mass balance considerations. For the LPE modification, equation (13) is applicable with $R_2 = 0$.

As an example of the use of equation (13), let the following be assumed for a sludge recycle system:

1. Plant flow equals 1,578 m^3/hr (10 mgd)

2. Minimum flow to stripper 10% of Q at 18,000 mg/l TSS

3. a equals 9.21×10^{-6}m/hr (30)

4. b equals 2.26 (30)

5. Elutriation efficiency greater than or equal to 0.50

6. Stripper retention time equals 10 hrs based on R_3

From Figure 9, f = 2.0 and the supernatant flow equals 1.5 times the sludge returned to aeration from the stripper so that $E \geq 0.5$ may be achieved. Initially this results in R_2 = 20% of Q and S = 6% of Q while R_3 = 4% of Q. A mass balance around the stripper indicates that C_3 = 45,000 mg/l. However, the level of C_3 is outside the range of predictability of equation (9). Therefore, a stripper underflow concentration of 25,000 mg/l is chosen to include a safety factor. Since 10% of Q is required to be fed to the stripper at 18,000 mg/l, then from a mass balance R_3 must equal .066 Q to achieve the same throughput. To achieve $E \geq 0.5$ with f = 2.0, S must now become 0.10 Q (Figure 9). Because $R_1 = R_3 + S$ and f = 2, R_1 becomes 0.166 Q while R_2 becomes .332 Q.

This set of conditions results in a reduction of the thickening capacity of the secondary clarifier because the solids flux in the stripper becomes controlling. Consequently, the blanket level in the secondary clarifier will become negligible and the underflow from the clarifier will average ~9900 mg/l. Under these assumptions, equation (13) predicts a cross-sectional area requirement of 2200 m^2 (23,690 ft^2).

If the low phosphate elutriation (LPE) system is employed, $R_2 = 0$ in equation (8), thus producing the minimum cross-sectional area. Further, since the supernatant flow is independent of the R_1 and R_3 streams in the LPE system, any desired level of supernatant flow may be achieved. Therefore, the cross-sectional area of the stripper and the lime requirements can be optimized independently for this modification.

For the LPE system, assumptions 1 through 6 are valid. It is also important that $C_1 = C_3$ so that all the supernatant formed in the stripper be a result of the elutriation and not thickening. (Thickening is to be avoided because thickening occurs before the phosphorus is released from the sludge and, as a consequence, the supernatant resulting from thickening contains very little phosphorus.) Therefore, $C_3 \approx 18,000$ mg/l and Figure 9 predicts a 10.5% of Q elutriant flow to produce an elutriation efficiency greater than 50%. This results in 10.5% Q as the supernatant. Applying equation (13) to the LPE system results in a stripper cross-sectional area requirement of 411 m^2 (4430 ft^3).

A comparison between these two options indicates that at approximately the same stripper supernatant rate - which is equivalent to the same chemical cost - the LPE modification requires less than 19% of the area of the sludge recycle system. The stripper volume for the LPE design would be 2079 m^3 while the volume of the sludge recycle unit would be 3735 m^3. Both volumes include 1.2 m of free liquid above the sludge blanket in the stripper. Under other conditions the difference between the two operating modifications may not be as great, and economical performance using the sludge recirculation mode can be achieved.

DESCRIPTION AND INTERACTION OF VARIABLES AFFECTING THE PROCESS

The performance of the process is highly dependent on the design of the stripper tank relative to the design of the activated sludge process. The interrelationships between the aerobic system and the anoxic sidestream will be established by reference to Figure 11. To evaluate the ability of the system to accumulate phosphorus within the biomass, two factors must be considered. First, the kinetics of the process must be understood, and second, the maximum capacity of the organisms with respect to phosphorus loading or capacity must not be approached.

The kinetics of phosphorus uptake has been shown to be zero order with respect to ortho phosphorus concentration down to very low levels. (25) The rate of phosphorus uptake is dependent on the general activity of the organisms in the aeration basin which may be related to food to microorganism ratio, F/M. Pilot scale testing has verified the importance of the F/M level to phosphorus uptake rate and has indicated a modest dependence of uptake kinetics on temperature. Therefore, the variables RT_A (based on total flow), BOD_5, and MLVSS are important as a result of the F/M relationship. Phosphorus removal rate generally increases with increases in F/M. The retention time and MLVSS parameters are also important to the kinetics of accumulation for another reason. Since the kinetics are zero order with respect to phosphorus, the amount of phosphorus taken up is directly proportional to the retention time available for uptake and the number of organisms taking up phosphorus (related to the MLVSS level). The number of organisms in the MLVSS that can take up phosphorus to form polyphosphate granules in turn depends on the mass of sludge entering and leaving the stripper tank. The number of organisms in the aeration tank possessing this uptake capability increases with increased sludge flow through the stripper. Thus, the PhoStrip process can be designed to take up any normal level of influent phosphorus by proper adjustment of the stripper mass throughput relative to RT_A. This adjustment will vary according to the quantity of phosphorus to be removed. Like other biochemical processes the kinetics of phosphorus removal can be related to temperature via an Arrhenius expression. Generally, the rate of removal increases with increasing temperature.

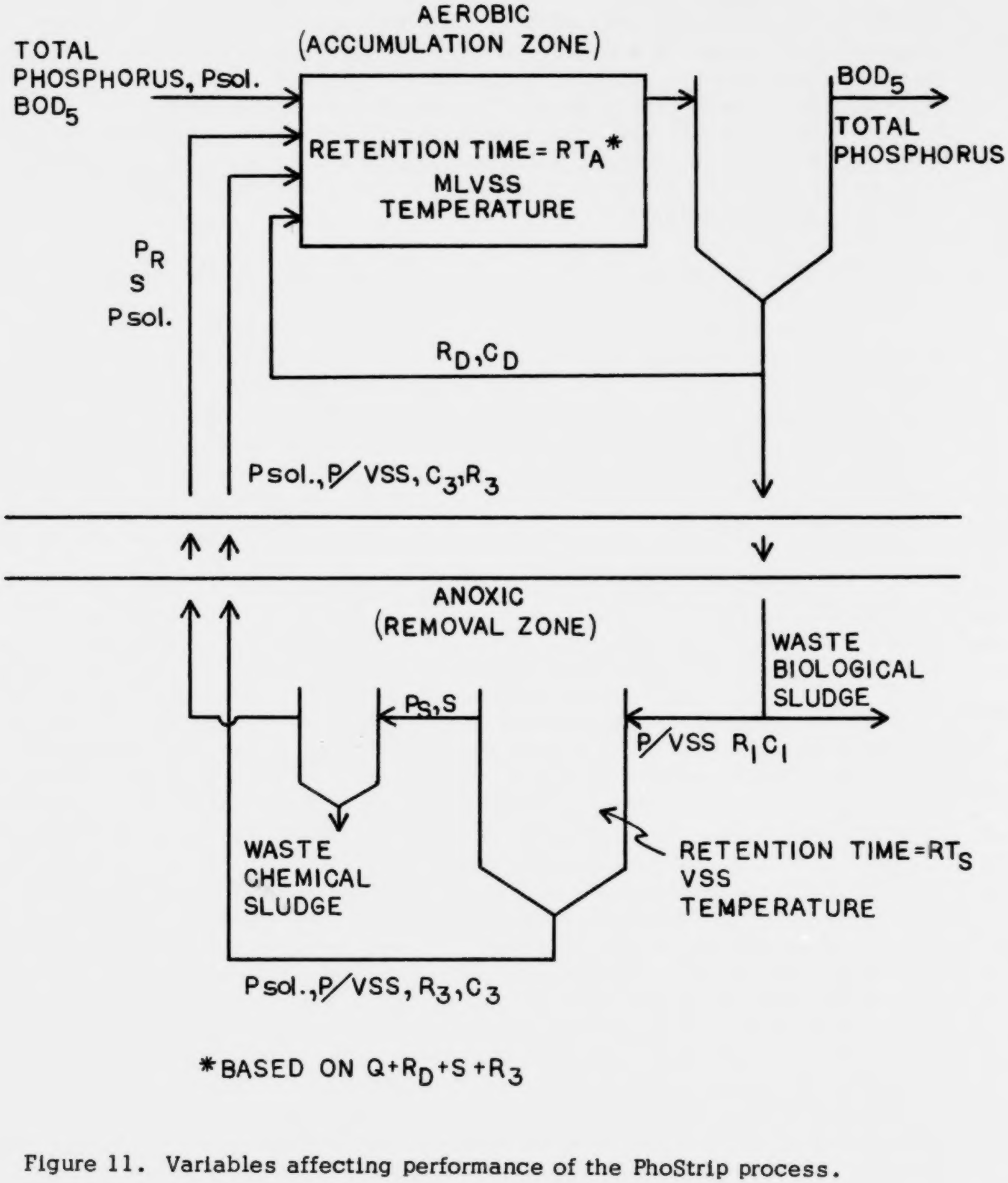


Figure 11. Variables affecting performance of the PhoStrip process.

From Figure 11 it can be seen that there are three sources of soluble phosphorus entering the aeration tank: that with the raw feed, that solubilized in the stripper but not appearing in the supernatant, and the phosphorus not precipitated via lime addition. The first factor is uncontrollable and may represent as much as 90% of the phosphorus to be taken up in the aeration basin. It is important to note that much of the non-soluble phosphorus in the influent is hydrolyzed by the activated sludge process and thus becomes soluble. The second factor, P_{sol}, of stream R_3 depends on the elutriation efficiency of the process. At low efficiencies, it is possible that more than 50% of the phosphorus entering the aeration basin may be solubilized from the polyphosphate granules. The last stream, the residual from lime precipitation, can be easily controlled so that it represents less than 5% of the influent to the aeration basin. Therefore, this stream may be ignored with respect to its contribution to phosphorus in the aeration basin.

The second major factor to be evaluated when considering the aerobic portion is that the ratio of the phosphorus level in the sludge to the phosphorus saturation level is dependent on many of the same parameters as the kinetics. The F/M in the aeration basin and mass flow through the stripper increase the ultimate capacity of the sludge for phosphorus by increasing the number of organisms with the potential of storing phosphorus. The F/M contributes to this increase by increasing the growth rate of the organisms. An increase in the mass flow through the stripper increases the capacity by stimulating more organisms to the metabolic pattern of phosphorus storage. The effect of F/M becomes apparent as an increase in the oxygen uptake rate of the sludge under aeration and as an increase in sludge wasting. The effect of mass flow through the stripper on phosphorus saturation level is not obvious. This latter fact is due to the effect of stripper performance on the P/VSS level of the sludge.

As in the evaluation of the aerobic zone, the performance of the anoxic zone with respect to phosphorus release depends on kinetics and the ratio of P/VSS in the sludge to the sludge's saturation level. Earlier studies (23) have indicated that the rate of phosphate release is dependent on the acetate level of the anoxic solution. This is consistent with Fuhs and Chen (17) who stated that a product was being produced by organisms not responsible for phosphorus uptake that facilitates release in the anoxic zone. The facultative organisms produce acetates and Krebs cycle intermediates during the anoxic period. The rate of formation of these compounds controls the rate of phosphorus release. Therefore, the length of time that a given mass of sludge is held in the stripper tank (RT_S) will affect the amount of acetate formed and thus phosphorus released. The rate of acetate formation per mass of sludge is also dependent on the overall activity of the sludge. An activated sludge operating at a high F/M will produce acetates more rapidly than a system at low F/M levels. Temperature also affects the rate of acetate formation which causes the release kinetics to exhibit an Arrhenius-type temperature relationship. Thus, the phosphorus level in the sludge leaving the stripper and entering the aeration basin can be controlled by the retention time in the stripper.

To optimize stripper performance, it is desirable to maximize phosphorus release per unit volume of stripper tank. This is accomplished by maintaining high suspended solids levels in the anoxic zone and optimizing the stripper sludge retention time. At very short retention times only small amounts of phosphorus are released from each unit mass of sludge put through the stripper. In order to solubilize enough phosphorus to achieve a low effluent phosphorus concentration, high stripper feed rates are required. Therefore, the size of the stripper tank must increase to handle the higher mass throughput rates. This mode of operation may increase the P/VSS level throughout the system depending on what fraction of the phosphorus removed ends up in the stripper supernatant (elutriation efficiency). High P/VSS levels could result in phosphate effluent quality problems due to the phosphorus content of the effluent volatile suspended solids.

On the other hand, the use of long anoxic periods with small sludge flow rates does not necessarily minimize stripper volume either. In order to utilize a low volume of sludge throughput, the sludge must contain large quantities of polyphosphate. From a microscopic view, the individual organisms entering the stripper will contain different levels of polyphosphate. Some organisms will have small concentrations of polyphosphates. When this level of polyphosphate is utilized, the organism can no longer release phosphorus, thus the rate of phosphorus release per unit mass of sludge decreases. Therefore, the phosphorus release per unit tank volume decreases. In order that sufficient phosphorus is released to achieve low soluble effluent phosphorus levels, a high P/VSS ratio must be attained in the aeration basin. However, long anoxic periods generally result in a small number of organisms with high individual phosphorus levels but low phosphorus levels on a P/VSS basis. Therefore, this procedure will also require high mass throughputs to achieve the overall phosphorus removal required. The flow rates under this mode will be substantially shorter than the short retention time mode but the long retention times can also result in high volumes. Also, the high P/VSS levels required can result in the same effluent quality limitations presented for the short retention time case.

Therefore, to minimize tankage requirements, a balance must be struck between retention time and mass throughput so that reasonable P/VSS levels are achieved in the sludge. The optimization of the stripper tank volume is highly dependent on elutriation efficiency. If a large fraction of the phosphorus that is released is washed into the supernatant, less phosphorus per unit mass needs to be released (mass balance) and the size of the stripper can be reduced.

At a given stripper size and mass throughput, the elutriation efficiency can be used to control the P/VSS level of the sludge according to the mass balance discussion presented earlier (Figure 6). The increase in chemical phosphate sludge mass via an increase in elutriation efficiency results in less phosphorus available to be removed with the waste activated sludge at the same sludge wasting rate. Therefore, the sludge will possess a lower P/VSS level. P/VSS can also be controlled via changes in throughput of sludge and stripper retention time. However, once the stripper tank is installed, the flexibility of this method of control is limited by the fixed size of the stripper tank. With the LPE system, control of the elutriation efficiency and thus P/VSS ratio is easily achieved by changing the elutriation rate while with the sludge recycle mode elutriation efficiency control is most easily achieved by controlling the stripper supernatant flow rate. This is accomplished by varying the stripper feed and sludge return rates.

Because of the effect of temperature on the kinetics of phosphate release the optimum retention time in the stripper tank will be influenced by the temperature of the wastewater.

SECTION 5

DESCRIPTION OF FACILITIES

MOBILE PILOT PLANT

The Union Carbide Mobile PhoStrip pilot plant was used to demonstrate the performance of a system designed to optimize the LPE system. Figure 12 is a schematic diagram of the mobile pilot plant which integrates a coarse bubble, diffused air activated sludge system with the PhoStrip phosphorus removal system. The nominal feed rate to the van is 0.11 m^3/hr. Although it is designed to optimize the performance of the LPE mode of operation, it is also capable of operating in the sludge recycle and stripper supernatant recycle modes of operation. However, the geometry of the tanks is such that less than optimum performance would be expected in the latter two modifications.

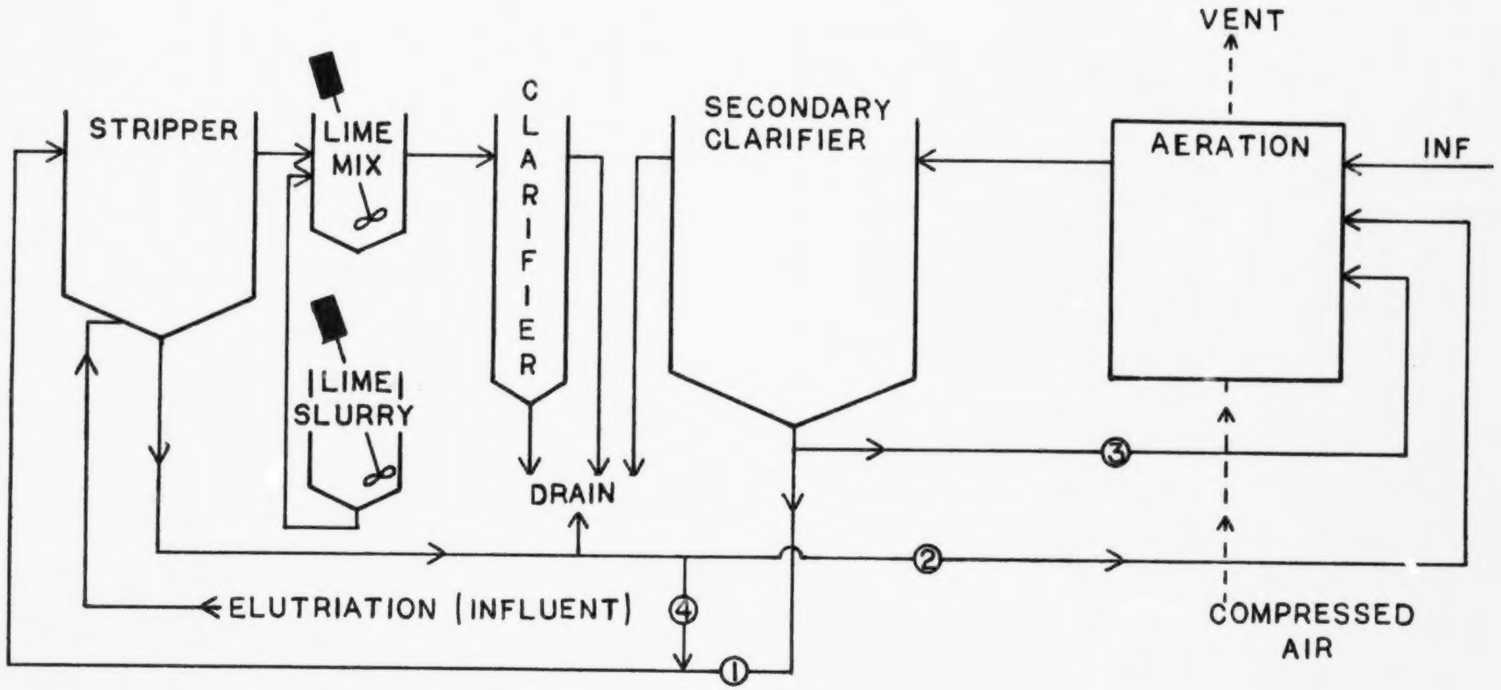
The basic components of the pilot plant as illustrated in Figure 12 are as follows:

1. Air Reactor: (Size: 2.44 m x 2.13 m x 1.37 m) The air reactor is constructed from stainless steel. It has removable baffles, an adjustable weir, and quick-connect couplings so that feed and recycle streams may be introduced at any desired location. Diffused air provides the mixing and mass transfer through Dravo leaf spring diffusers. The reactor has the capability of being operated as a completely mixed system, a conventional activated sludge system, and a step feed or contact stabilization unit.

Two Dravo diffusers are situated in each of the three passes of the serpentine type conventional activated sludge system. The compressed air is provided by six Model 4907 Thomas compressors.

2. Secondary Clarifier: (Size: 1.22 m diameter x 1.37 m side wall depth) The secondary clarifier is a center feed, peripheral overflow type with plow-type scrapers. The sludge is removed through a discharge line at the center of a conical bottom. Part of the sludge can be pumped to the stripper tank while the remainder may be returned directly to the aeration basin.

3. Stripper Tank: (Size: 1.12 m diameter x 1.83 m side wall depth) The stripper tank is a standard gravity-type thickener with center feed,



① STRIPPER FEED
② STRIPRER SLUDGE RETURN
③ SECONDARY SLUDGE RETURN
④ SLUDGE RECYLE OPTION

Figure 12. Schematic Diagram - PhoStrip Pilot Plant, Reno/Sparks, Nevada.

peripheral overflow and center sludge withdrawal. No rakes were added to induce thickening. A distribution arm for use with the LPE and stripper supernatant recycle systems is located at the bottom of the tank. The arm is feed liquid from a pump through a rotating unit at the top of the tank. The sludge recycle mode utilizes a pump to remove sludge from the bottom of the stripper and return it to the top of the tank through the stripper feed line. A second sludge pump is used to return sludge from the bottom of the stripper to the aeration basin.

4. Lime Mix Tank: (Size: 0.61 m diameter x 0.91 m deep) The lime mix tank receives phosphorus-rich supernatant liquid from the stripper overflow and lime slurry from the lime slurry tank. Both are thoroughly mixed by a mixer mounted on top. Any desired pH above that of the stripper supernatant and below that of the lime slurry (pH ~12.5) may be achieved.

5. Lime Slurry Tank: (Size: 0.61 m diameter x 0.91 m deep) The lime slurry tank contains the lime which is pumped to the mix tank.

6. Clarifier: (Size: 0.46 m diameter x 1.37 m depth) This clarifier receives the overflow from the lime mix tank. The clarified liquid is either pumped back to the aeration basin or it may be used as an elutriant in the LPE system. The phosphate containing chemical sludge underflow is pumped to the drain.

RENO/SPARKS WATER POLLUTION CONTROL PLANT

The Reno/Sparks Water Pollution Control Plant is a jointly-owned facility serving the cities of Reno and Sparks, Nevada. The plant is designed for an average flow of 76,000 m^3/d (20 mgd) and discharges into the Truckee River. A simplified flow schematic of the facility is presented in Figure 13. The pump station collects the wastewater from the two municipalities and feeds the treatment plant through variable speed controlled pumps. The wastewater passes through an aerated grit chamber to the primary and secondary treatment systems which are modularized. There are three modules, each consisting of a primary clarifier, an aeration basin, and a final clarifier. The capacity of each module is 25,000 m^3/d (6.67 mgd).

The primary clarifiers are square units 25.6 m across by 3.3 m side water depth. These are center feed units with peripheral overflow. Plow-type scrapers on a modified mechanism similar to that used in circular clarifiers were utilized to transfer the sludge from the floor of the clarifiers to a central withdrawal well.

The aeration basins are diffused air units with coarse bubble diffusers. Each modular aeration basin employs a three-pass serpentine design, each pass having the approximate dimensions 54.9 m x 8.2 m x 5.3 m side wall

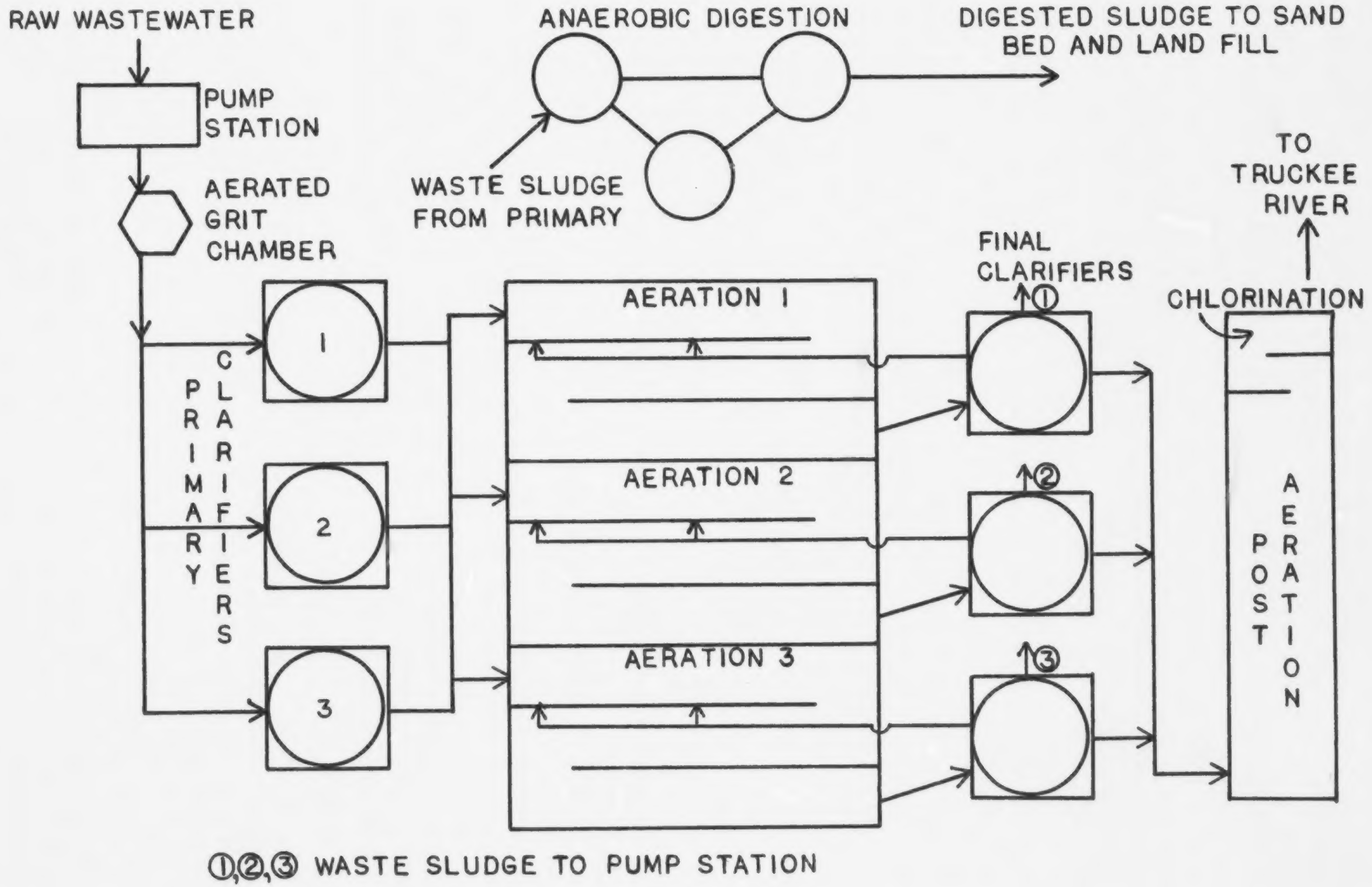


Figure 13. General flow schematic - Reno/Sparks Water Pollution Control Plant.

depth. The plant has the capability of adjusting the aeration rate to correspond to the oxygen demand at various locations.

The secondary clarifiers are square tanks 25.6 m across and 3.3 m side water depth. The secondary clarifiers use rapid sludge return (RSR) siphon units to pick up sludge for return to the aeration tank. As was the case in the primary system, the sludge transfer unit was the type normally employed in circular clarifiers with modifications to accommodate use in square tanks. The sludge pickup units deposit sludge in a wet well from which it is pumped to the aeration basin. The return sludge pumps are variable speed and discharge at two points in the aeration basin, one at the head end and one midway through the first pass of each module. Venturi meters are used to measure return sludge flow rates.

The effluent from the secondary clarifier flows to a post aeration basin to increase the dissolved oxygen concentration to a level above that required for water in the Truckee River (D.O. $>$ 6.0 mg/l). Following post aeration the effluent is chlorinated and then discharged.

The waste sludge from the aeration basins flows by gravity to the wet well of the raw sewage pump station. This material settles out in the primary clarifier along with the primary solids. The combined primary plus secondary sludge is periodically pumped from the primary clarifiers to a two-stage anaerobic digestion process. The digested sludge is dewatered on sand beds and trucked to a land fill.

MODIFICATIONS OF THE FULL SCALE PLANT

The Reno/Sparks Joint Water Pollution Control Plant was modified twice for evaluations of the PhoStrip process. The first modification was initiated by Kennedy Engineers, 657 Howard Street, San Francisco, California 94105, to test the general performance of the process utilizing the sludge recycle option. (31) After testing on the sludge recycle mode was completed, the flow configuration of the process was changed to that of the LPE system.

During both of the test periods, the primary clarifier labelled #1 in Figure 13 was utilized as the stripper tank; and, as a result, Primaries #2 and #3 were operated at approximately 150% of design load because the flow to the plant was near design capacity. The effluent from these two primary clarifiers was equally distributed to the three aeration basins.

SLUDGE RECYCLE MODIFICATION

During this phase of operation, the secondary aeration basin labelled #1 was used to assess the performance of the PhoStrip system. Thus, 1/3 of the total plant flow (25,000 m^3/d) was treated via the PhoStrip process.

Aeration basins #2 and #3 were used as controls to determine comparative removal efficiencies for conventional activated sludge systems. In order to achieve this condition, primary clarifier #1 and aeration basin #1 were modified according to Figure 14. The return sludge piping was extended from the head end of the aeration basin to the distribution unit near the center of the stripper tank. The 406 mm valves in the return sludge line were used to manipulate the flow to the stripper tank and aeration basins. Because the variable speed recycle pump control system would not function properly below a minimum flow rate, the 406 mm valve midway down the first aeration pass was used to maintain at least this minimum flow from the secondary clarifier. The distribution device near the center of the stripper consisted of a cylindrical tube baffled at the bottom by a cone. The cone imparted some horizontal velocity to the sludge which left the cylinder 1.8 m below the surface of the liquid. A pump was installed to remove the sludge from the bottom of the stripper tank for return to the aeration basin and to provide the sludge recycle stream. The output of the pump was calibrated and the relative sludge recycle and return flows were obtained from determinations of relative flow rates as a function of the settings on the 152 mm valve that discharged to the aeration basin. The anoxic recycle sludge was pumped to the central distribution device. The supernatant from the stripper tank was siphoned to a drainage ditch where it flowed to a juncture with the chlorinated plant effluent. The supernatant flow rate was determined from calibrations made previously on the 152 mm "Ballcentric" type plug valve.

LPE SYSTEM MODIFICATION

As was shown in the example presented with the Solids Flux Theory, the use of the LPE system reduces the cross-sectional area requirement relative to that of the sludge recycle system. Since the primary clarifier (stripper tank) at the Reno/Sparks plant had a fixed cross-sectional area, the solids flux considerations indicated that two-thirds of the plant flow (50,000 m^3/d) could be treated with the available area. As a result, the system was modified according to the flow schematic presented in Figure 15. Aeration modules #1 and #2 were converted to the PhoStrip process. The existing sludge return pumps were used to pump settled activated sludge to the head end of each module's aeration basin. At that point, the sludge streams were split. Some of the sludge was discharged through a 152 mm valve to the aeration basin while the remainder continued on to the center well of the stripper. No special inlet device was required during this mode of operation. The relative flows to the stripper and aeration basins were obtained from measurements of the total recycle flow with the Venturi meters and the stripper feed flows via magnetic flow meters. The return sludge from the stripper was transferred to the aeration basin utilizing the same pump as that employed in the sludge recycle system tests. The discharge from the pump was split into two streams which emptied into the head end of each of the aeration basins. The flow of these streams was determined through the use of magnetic flow meters.

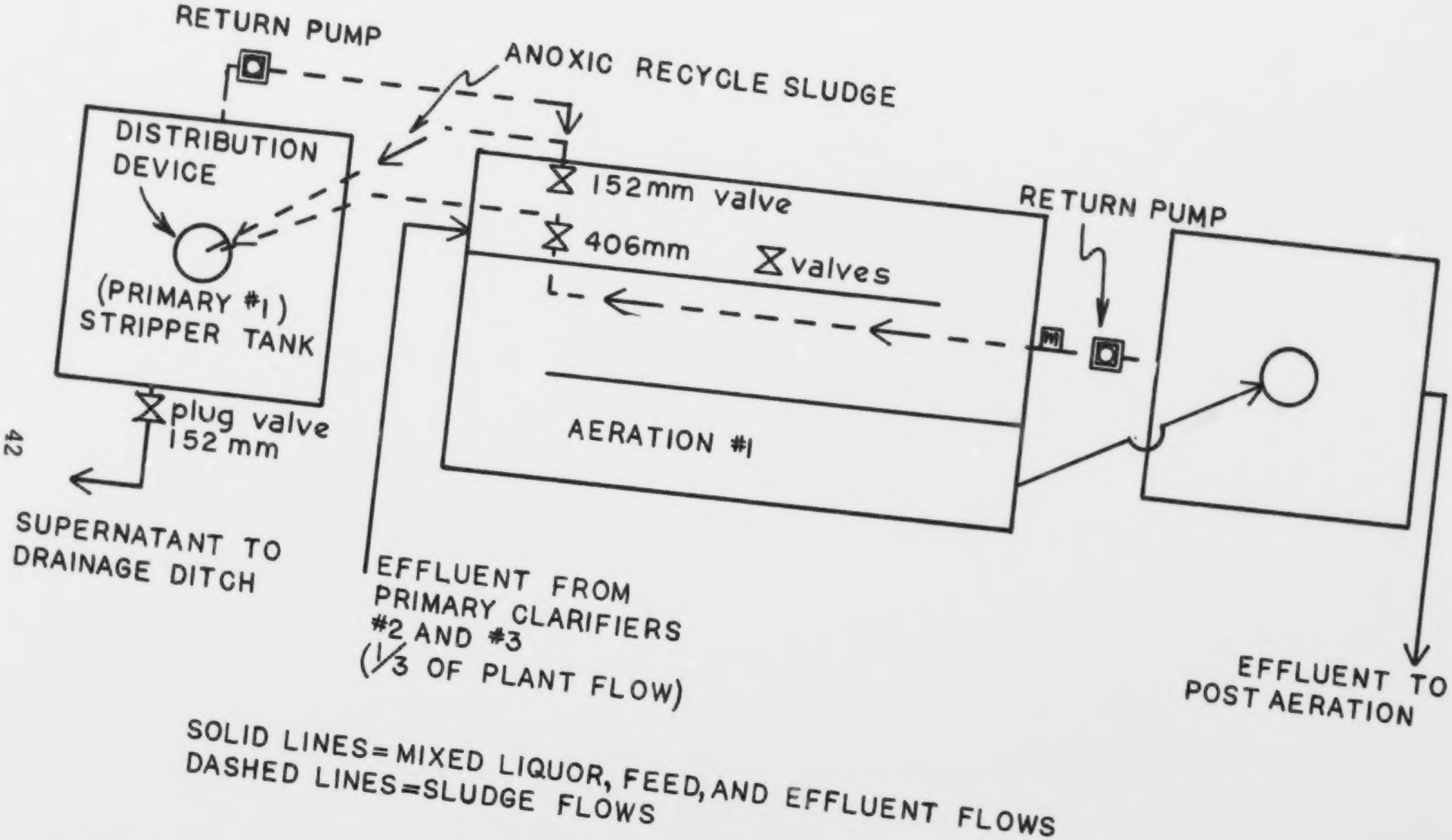

RETURN PUMP
ANOXIC RECYCLE SLUDGE
DISTRIBUTION DEVICE
(PRIMARY #1) STRIPPER TANK
152mm valve
406mm
valves
RETURN PUMP
plug valve 152 mm
AERATION #1
SUPERNATANT TO DRAINAGE DITCH
EFFLUENT FROM PRIMARY CLARIFIERS #2 AND #3 (1/3 OF PLANT FLOW)
EFFLUENT TO POST AERATION
SOLID LINES=MIXED LIQUOR, FEED, AND EFFLUENT FLOWS
DASHED LINES=SLUDGE FLOWS

Figure 14. Reno/Sparks Water Pollution Control Plant - PhoStrip system flow diagram - Sludge recycle system.

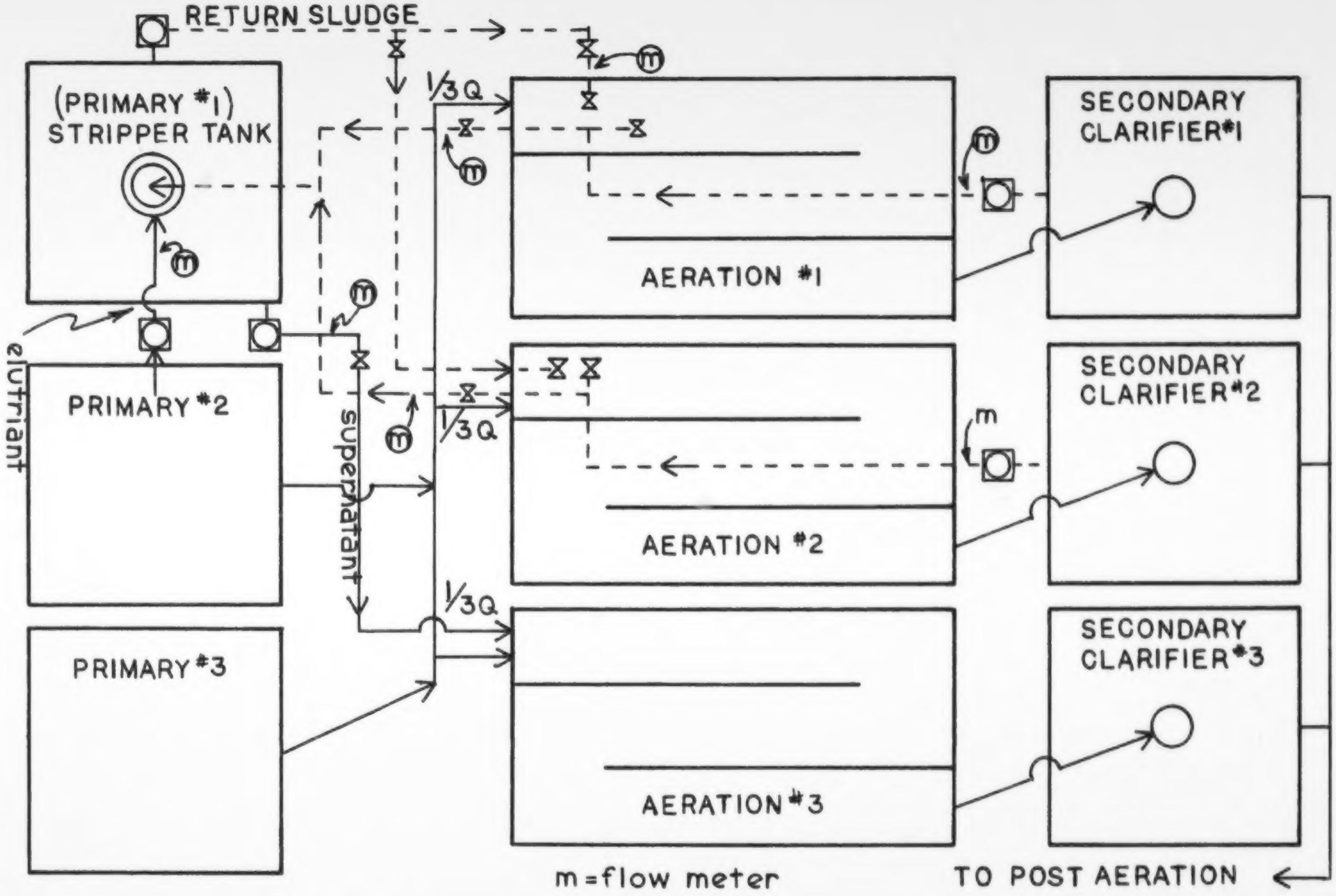


Figure 15. Reno/Sparks Water Pollution Control Plant - PhoStrip system control diagram - Low phosphate elutriation system.

The elutriant employed during the period of this study was the primary effluent from primary module #2. This LPE stream was pumped to a distribution system. The top of the system was a circular basin divided into four concentric rings, each being fed elutriant individually. Each concentric ring fed a vertical pipe that transferred the elutriant to a horizontal arm. Two of these arms distributed the elutriant over the inner half of the tank and two arms were used to cover the outer area. The arms were supported from the truss bridge at the lowest possible location. The entire distribution system, including basin and arms, rotated along with the scraper mechanism. The elutriant flow was measured via a vortex shedding meter.

The supernatant from the stripper was pumped to aeration basin #3 so that no unchlorinated material would enter the Truckee River. The flow rate of this stream was measured through the use of a second vortex shedding meter.

SECTION 6

RESULTS & DISCUSSION

PILOT SCALE EVALUATION

The pilot scale program performed at Reno/Sparks, Nevada was divided into three phases of operation. The first phase employed the sludge recycle system to produce a stripper supernatant high in soluble phosphorus. The second and third phases used the LPE modification with primary clarifier effluent employed as the elutriant. Phases II and III differ in that the stripper supernatant was not treated during Phase II, while during Phase III the supernatant was treated with lime and returned to the aeration basin.

The procedure used to evaluate the performance of the pilot plant was to set up the desired operating conditions on the pilot plant and monitor only a few basic parameters: flow rates, suspended solids, and orthophosphates; until the plant stabilized and achieved steady state operation. Once steady state was obtained, the system was intensively monitored for a period of from two to three weeks. Table 5 presents the analytical tests performed during the periods of intensive evaluation. Samples for analysis of influent, effluent, and stripper supernatant were 24-hour composite samples. All other samples were grabbed once each day. All analyses were performed according to U. S. EPA (32) approved procedures. The process operation monitoring was performed once each day and included measuring and adjusting all flow rates (influent, elutriation, stripper influent and underflow, and lime slurry feed); measuring blanket levels in the secondary clarifier, stripper, and chemical clarifier, and taking reactor temperature and dissolved oxygen readings. Wasting of sludge from the secondary clarifier was performed both manually and automatically. Wasting of the lime precipitated phosphorus sludge was performed manually to maintain a constant blanket level in the chemical clarifier.

The pilot plant arrived at the Reno/Sparks Water Pollution Control Plant on October 15, 1975. The system was seeded with recycle sludge from the existing full scale plant during the first week of November. Since this was the first time this pilot plant had ever been operated, several mechanical problems were encountered which had to be solved during the course of the program. Following startup and shakedown, an intensive sampling and analytical program was performed from November 19, 1975 through December 8, 1975,

TABLE 5. ANALYTICAL MONITORING OF THE PHOSTRIP PILOT PLANT, RENO/SPARKS, NEVADA

Parameter	Frequency	Sample Point
1. COD Total	Daily	Influent Effluent
2. TSS & VSS	Daily	Influent Effluent Secondary Clarifier Underflow Stripper Underflow Stripper Supernate Mixed Liquor
3. Total P	Daily	Influent Effluent Stripper Supernate Aeration Effluent Filtered Aeration Effluent Secondary Clarifier Underflow Filtered Clarifier Underflow Stripper Underflow Filtered Stripper Underflow Chemical Clarifier Supernatant
4. Ortho P	Daily	Influent Effluent Stripper Supernate

Phase I. Primary settled wastewater from the full scale plant was used as feed for the pilot plant. The supernatant from the stripper was allowed to flow to drain. As will be discussed later, the final effluent quality from this mode of operation was not as good as anticipated and the next month was spent trying to improve the performance of the sludge recycle mode. On January 19, 1976 the pilot plant was switched to operation in the LPE mode with primary clarified effluent as the elutriant. After mechanical problems were solved and steady state operation was achieved, the Phase II intensive evaluation began. Phase II lasted from February 23, 1976 through March 7, 1976. During this period the supernatant from the chemical clarifier was allowed to go to drain. Therefore, Phase III was initiated on March 8, 1976 to run the system in such a manner as to remove phosphorus from the system only through wasting of the biological or chemical sludges. The period of intensive evaluation lasted from March 20 through April 2, 1976.

Table 6 contains a summary of the operating conditions maintained during the intensive evaluation period of Phase I. The conditions maintained on the pilot plant are indicative of average operating conditions observed for the full scale plant. The flow rate and suspended solids data around the stripper are typical of an air PhoStrip system which produces a supernatant through thickening. The f parameter is used for a comparison of the actual elutriation efficiency for this system with the theoretical model presented earlier. The food to mass ratio (F/M_A) presented in Table 6 translates to an F/M_A of approximately 0.45 on a BOD_5 basis. This loading rate is typical of the Reno/Sparks full scale secondary system.

Table 7 contains a phase average summary of the pertinent analytical data taken during this period of operation. The data on effluent COD and suspended solids shows only fair performance. This is a result of the mechanical problems encountered during this first phase of operation on the mobile pilot plant. The effluent quality data (appendix) shows a general trend toward improvement with time during Phase I. The total phosphorus data indicates that only a 56% reduction in the influent phosphorus level was achieved across the system. A close look at the data indicates that very little of the phosphorus released by the organisms in the stripper appeared in the supernatant. This low elutriation efficiency ($\epsilon = 0.15$) is responsible for a high phosphorus loading on the aeration basin (F_P/M_R). The P/VSS level in the clarifier underflow suggests that the uptake of phosphorus in the aeration basin was kinetically limited since values higher than 0.044 have been achieved with the PhoStrip process.

The clarifier underflow P/VSS ratio also indicates that approximately 1.3 mg P/l of effluent was a result of effluent suspended solids. This conclusion is in variance with the difference between the effluent total and ortho phosphorus levels. There are three possible contributory factors that cause this apparent contradiction. First, it is possible that the organisms that

TABLE 6. PHASE I OPERATING CONDITIONS AT THE PHOSTRIP PILOT PLANT, RENO/SPARKS, NEVADA
(SLUDGE RECYCLE, 11/19-12/8/75)

Feed Flow Rate, Q (m^3/day)	27.4	
Stripper Feed Rate, R_1 (m^3/day)	6.7	
Stripper Underflow Rate, R_3 (m^3/day)	3.4	
Stripper Supernatant Rate, S (m^3/day)	3.4	(12.3% of Q)
Sludge Recycle Rate (around stripper), R_2 (m^3/day)	2.9	($f = \frac{R_2}{R_1} = .43$)
Waste Activated Sludge Rate (m^3/day)	0.31	
Retention Time Aeration (based on Q), hr.	4.9	
Retention Time Anoxic (based on R_3), hr.	8.0	
MLSS, mg/l	1370	
MLVSS, mg/l	1200	
R_1SS, mg/l	6380	
R_1VSS, mg/l	5470	
R_3SS, mg/l	12,390	
R_3VSS, mg/l	10,260	
TSS, Stripper Supernatant, mg/l	48	
VSS, Stripper Supernatant, mg/l	43	
Dissolved Oxygen, Mixed Liquor (mg/l)	1.8	
pH, Mixed Liquor	6.9	
Temperature, °C	15.6	
F/M_A (kg COD/day/kg MLVSS)	0.93	
Stripper Blanket Level, m.	1.2	
Secondary Clarifier Blanket Level, m.	0.23	

TABLE 7. PHASE I AVERAGE ANALYTICAL RESULTS AT THE PHOSTRIP PILOT PLANT, RENO/SPARKS, NEVADA
(SLUDGE RECYCLE, 11/19-12/8/75)

COD_{total}, mg/l	
Influent	228
Effluent	47
Total Suspended Solids, mg/l	
Influent	126
Effluent	36
Effluent Volatile Suspended Solids, mg/l	30
Total Phosphorus, mg P/l	
Influent	8.0
Effluent (ortho)	3.5 (3.2)
Stripper Supernatant	13.6
Stripper Underflow	413
Filtered Stripper Underflow	77
Filtered Aeration Effluent	2.6
Secondary Clarifier Underflow	242
Secondary Clarifier Underflow Filtered	12
P/VSS	
Secondary Clarifier Underflow	0.044
Stripper Underflow	0.033

concentrate the phosphorus do not become equally distributed in the effluent suspended solids and the clarifier underflow. According to Fuhs (17), organisms of the genus Acinetobacter tend to form clumps which would improve their settling characteristics. Second, the phosphorus analysis of the sludge samples are grab samples. The level of phosphorus in the sample depends on the conditions in the sludge during the period of sampling. Large variations can occur as a result of the diurnal variations in influent phosphorus levels. Finally, the effluent samples are 24-hour composites. As a consequence, the suspended solids were maintained in an anoxic condition for up to 24 hours without refrigeration. This set of conditions could be responsible for the resolubilization of significant quantities of intracellular polyphosphates. The difference between the observed and calculated effect of suspended solids on the effluent quality is believed to be mainly a result of a combination of arguments two and three.

Further proof of the effect of grab samples appears in the determination of phosphorus released in the stripper tank. A phosphorus mass balance around the stripper using the change in P/VSS ratio to determine phosphorus release results in a 63% higher level of phosphate solubilization than that obtained by calculations made on the supernatant composite and the stripper underflow filtered grab sample. In this case, the underflow grab sample from the stripper approaches the equivalent of an 8-hour composite because of the sludge recirculation mode employed while the clarifier underflow grab sample is an instantaneous sample resulting from the pseudo plug flow nature of the aeration basin. Therefore, the second method of calculating phosphorus resolubilization is more accurate. This indicates that the average P/VSS level of the sludge leaving the secondary clarifier averaged significantly less than the measured value.

The marginal difference between the clarifier underflow filtered phosphorus concentration and the stripper supernatant concentration on Table 7 may again be partially a result of comparing grab to composite samples. However, the low elutriation efficiency ($\varepsilon = 0.15$) compared to that predicted theoretically from Figure 8 ($\varepsilon \not\approx 0.23$) tends to indicate that the release of phosphorus in the sludge was inhibited (possibly by dissolved oxygen) in the upper portion of the stripper tank. The effect of dissolved oxygen would be to cause a rapid uptake of phosphorus by the organisms, thus reducing the quantity appearing in the supernatant. The performance during Phase I is consistent with the performance of the full scale plant at Reno/Sparks, Nevada before the specially modified central distribution device was added to the system. (33)

A summary of the operating conditions for Phases II and III is presented in Table 8. The conditions described in this table are only slightly different from those of Phase I. Of significance is the reduction in the amount of thickening which occurs in the stripper tank. A comparison of the elutriation and

TABLE 8. PHASES II AND III OPERATING CONDITIONS AT PHOSTRIP PILOT PLANT, RENO/SPARKS, NEVADA

	Phase II LPE (2/23-3/7/76)	Phase III LPE & Precipitation (3/20-4/2/76)
Feed Flow Rate, Q (m^3/day)	25.7	28.1
Stripper Feed Rate, R_1 (m^3/day)	3.8	3.6
Stripper Underflow Rate, R_3 (m^3/day)	2.9	3.1
Stripper Supernatant Rate, S (m^3/day)	4.8	3.1
Elutriation Rate (m^3/day)	4.1	2.6
Waste Activated Sludge Rate (m^3/day)	0.39	0.17
Retention Time Aeration (based on Q)	6.0	5.45
Retention Time Anoxic (based on R_3)	12.5	11.4
MLSS, mg/l	960	1070
MLVSS, mg/l	870	950
R_1SS, mg/l	5,390	8,070
R_1VSS, mg/l	4,500	7,010
R_3SS, mg/l	7,730	10,540
R_3VSS, mg/l	6,550	8,970
TSS, Stripper Supernatant, mg/l	44	34
VSS, Stripper Supernatant, mg/l	37	32
Dissolved Oxygen, Mixed Liquor (mg/l)	2.0	1.6
pH, Mixed Liquor/Chemical Clarifier	7.1	7.3/9.4
Temperature, °C	15.	16.7
F/M_A (kg COD/day/kg MLVSS)	1.08	1.12
Stripper Blanket Level, m.	1.6	1.5
Secondary Clarifier Blanket Level, m.	0.11	0.15

the stripper supernatant flow rates indicates that only 15% of the supernatant flow was a result of thickening. In Phase I, 100% of the supernatant was produced through thickening. A significant addition to this table is the pH level at which the phosphate precipitation was performed during Phase III. Since this phase is the only one which utilized a fully closed loop system, the importance of the pH level of precipitation becomes apparent as an important factor affecting the overall performance of the process.

Table 9 contains a summary of the analytical results obtained during these two periods of operation. The effluent COD and suspended solids level improved substantially over that observed earlier. This was mainly due to an improved understanding of the operational characteristics of the pilot plant. The effluent quality of the system with respect to phosphorus also shows a significant improvement over Phase I. A closer examination of the data indicates that the improved effluent quality resulted from the higher elutriation efficiency (ε). In Phase II, approximately 71% of the phosphorus released in the stripper was washed into the supernatant stream, while in Phase III the efficiency approached 58%. These elutriation efficiencies reduced the phosphorus loading to 0.054 kg P/day/kg MLVSS in Phase II and 0.067 kg P/day/kg MLVSS in Phase III. The previously established maximum removal rate for this set of conditions was ~0.06 kg P/day/kg MLVSS from Phase I. Since the rate of application of phosphorus in Phase II was below this number, consistent effluent quality was achieved shortly after the system was stabilized. Figure 16 demonstrates the stability of the effluent quality for this period of operation. Figure 16 also shows the fluctuation of effluent quality in Phase III. The F_P/M_A parameter calculated on an average basis was slightly higher than the maximum F_P/M_R permitted by the system. Thus, an increase in the phosphorus loading due to reduced elutriation efficiency or an increase in the phosphorus feed level results in an increase in effluent phosphorus concentration. On March 21-23, 1976, problems were encountered in the performance of the lime addition system. The addition of lime was hindered by erratic performance of the lime slurry feed pump resulting in the inability to maintain the pH in the quick mix tank above 8.5. Therefore, the supernatant ortho phosphorus concentration rose to a high of 39 mg/l on March 23 when the pH was measured at 7.7. This lack of pH control resulted in an increase in the F_P/M_A parameter to a point substantially higher than the 0.064 average maximum removal rate. Consequently, the system was kinetically limited as the increase in effluent phosphorus levels indicates. An improvement in the ability to control pH was achieved on the 24th, and the effluent quality improved accordingly. However, on the 26th the maximum removal rate was again significantly exceeded by the phosphorus loading. This time the increased load was due to an increase in the feed phosphorus concentration. The increase was not as great as in the previous incident, and, as a result, the effluent response was not as pronounced.

The performance of Phase III could have been improved by increasing

TABLE 9. PHASES II AND III AVERAGE ANALYTICAL RESULTS AT PHOSTRIP PILOT PLANT, RENO/SPARKS, NEVADA

	Phase II LPE (2/23-3/7/76)	Phase III LPE & Precipitation (3/20-4/2/76)
COD_{total}, mg/l		
Influent	233	241
Effluent	26	29
Total Suspended Solids, mg/l		
Influent	93	98
Effluent	17	27
Effluent Volatile Suspended Solids, mg/l	14	21
Total Phosphorus, mg P/l		
Influent	8.4	9.4
Effluent (ortho)	0.7 (0.5)	0.8 (0.5)
Stripper Supernatant	33	56
Stripper Underflow	216	386
Filtered Stripper Underflow	30	45
Filtered Aeration Effluent	0.8	0.8
Secondary Clarifier Underflow	190	338
Chemical Clarifier Overflow (ortho)	6.7 (3.5)	11.3 (2.4)
P/VSS		
Secondary Clarifier Underflow	0.042	0.048
Stripper Underflow	0.028	0.038

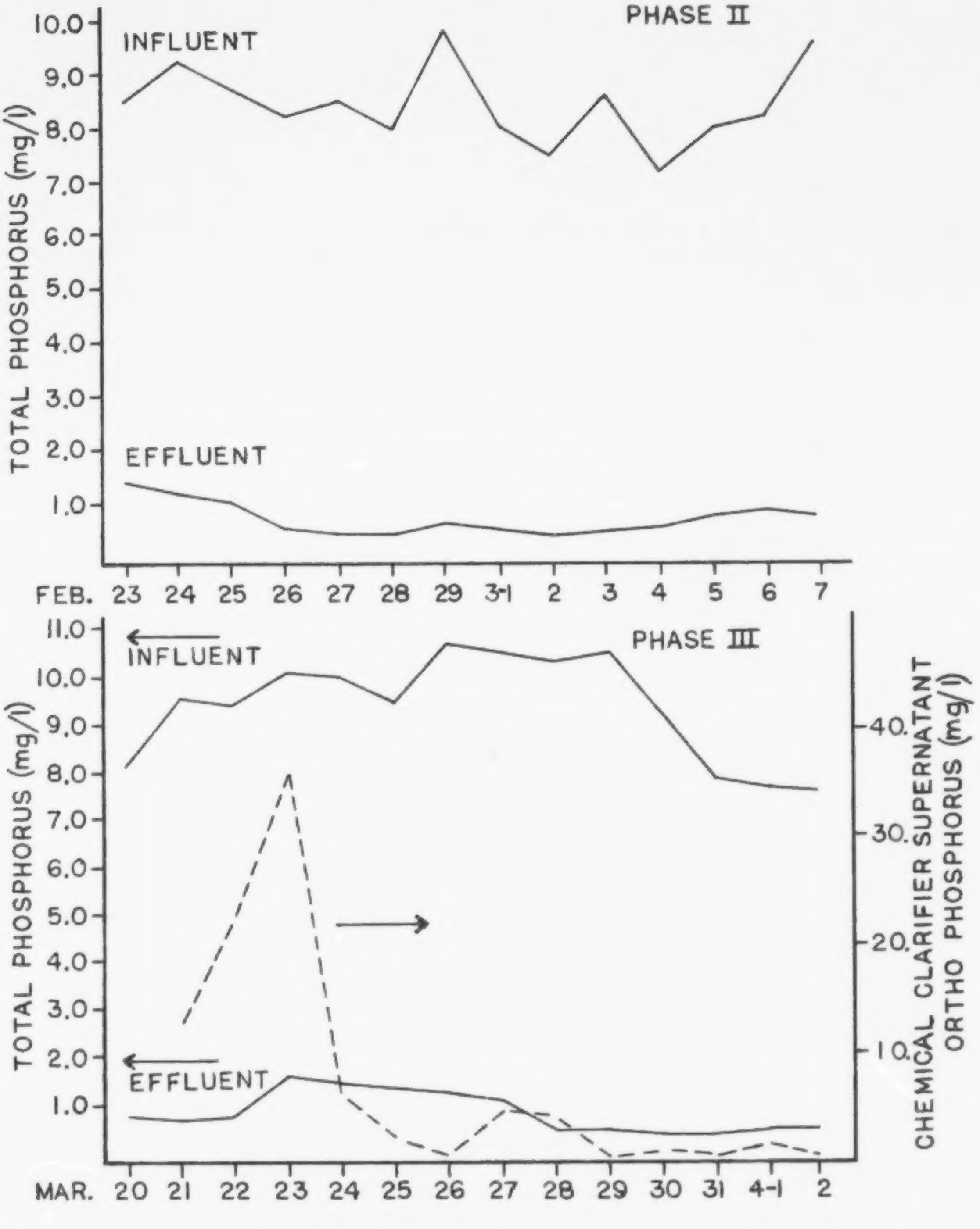


Figure 16. Daily influent and effluent phosphorus concentrations. PhoStrip pilot plant, Reno/Sparks, Nevada.

the MLVSS level, increasing the aerobic retention time or improving the elutriation efficiency. In Phase III, an "ε" value equal to 0.58 was obtained while in Phase II, $\varepsilon = 0.71$. The use of Figure 9 shows that the elutriation efficiency could have been improved in Phase III by increasing the elutriation rate. When comparing the ε data in these phases to the LPE System curve of Figure 9, it is obvious that the LPE curve is conservative relative to the performance of the pilot plant.

Table 9 shows that the contribution of phosphorus to the effluent due to suspended solids is less than that anticipated by the P/VSS levels in the clarifier underflow. Again, this is probably due to the combination of reasons presented in the discussion of Phase I.

Table 10 summarizes the phosphorus mass balances performed around the system for each period of intensive evaluation. Phase I shows only moderate phosphorus removal due to the low fraction of phosphorus removed with the stripper supernatant stream. In each phase the biological sludge wasting was shown to be a significant contributor to the overall phosphorus removed by the process. However, Phase I demonstrates that the kinetics of phosphorus uptake prevents activated sludge wasting from efficiently removing a high fraction of phosphorus from the influent. The increased elutriation efficiency is mainly responsible for the improved performance of the last two phases since no significant increase in phosphorus removed through sludge wasting was observed over that in Phase I. The data for stripper supernatant P related to Phase III represents only the phosphorus removed through chemical precipitation. To determine this quantity the total phosphorus of the chemical clarifier supernatant that was returned to the aeration basin was subtracted from the total phosphorus of the stripper supernatant. The errors in the balances are a result of procedural and analytical inaccuracies. However, the magnitude of the errors are not sufficient to invalidate the conclusions which may be drawn. The largest error exists in the determination of the accumulation term because of the necessity for using grab samples. The accumulation term is a measure of the change in phosphorus inventory in the aeration basin, the secondary clarifier, and the stripper tank.

From the point of view of the process designer these data suggest that the overall F_P/M_A be maintained at less than 0.06 kg P/day/kg MLVSS with an anoxic retention time of between 5.5 and 15 hours. Further, between 80 and 100 kg of VSS should enter the stripper tank for each kilogram of phosphorus in the raw feed. The elutriation efficiency levels which are necessary to achieve 90% plus reduction of influent phosphorus levels will be determined by the mass balance considerations. Generally, the required efficiency increases with increased influent phosphorus level. The performance of the sludge recycle system indicates that this system employing the conventional centerwell distribution systems results in poor phosphorus elutriation possibly due to inhibition of phosphorus release by the organisms in the upper portion of the

TABLE 10. PHASE AVERAGE PHOSPHORUS BALANCE DATA AT PHOSTRIP PILOT PLANT, RENO/SPARKS, NEVADA

	Phase I	Phase II	Phase III
Influent P, g/day	219	250*	288*
Effluent P, g/day	96	18.0	24.6
% of Influent	43.7	7.2	8.5
Stripper Supernatant P, g/day	46	158	175
% of Influent	20.9	63.2	60.7
Waste Activated Sludge Plus Accumulated P, g/day	81.6	71	62.5
% of Influent	37.3	28.4	21.7
P Balance Error, % of Influent	+ 2.1	- 1.2	- 8.9

* Based on influent plus elutriation flow

blanket. This inhibition may be due to the entrapment of small quantities of dissolved oxygen with the sludge as it enters the stripper tank. The dissolved oxygen may be a result of not being able to maintain a blanket level in the secondary clarifier when sludge recycle is employed. Diffusion of oxygen through the air liquid interface of the stripper may also be partially responsible. The data also shows that the LPE system avoids these problems and that the elutriation can be controlled by controlling the flow rate of the elutriant relative to the stripper underflow rate. The "ε" curve presented in Figure 9 is conservative relative to the LPE system's performance on this pilot plant. The performance of the process was also shown to be a function of the efficiency of the phosphate precipitation. The efficiency is controlled by the pH of precipitation. A target pH of 9.4 was seen to be sufficient to produce an ortho phosphorus level of less than 1.0 mg P/l in the chemical clarifier supernatant.

FULL SCALE TESTING OF THE SLUDGE RECYCLE SYSTEM

The initial evaluation of the PhoStrip process at Reno was performed by Kennedy Engineers, San Francisco, California, under the direction of L. E. Peirano. This work began in May 1974. Incomplete phosphorus removal was obtained until mid-September 1974, when a number of plant modifications were incorporated into the system. Of major importance was the modification of the inlet to the stripper as described in Section 5. From that point, the Reno/Sparks PhoStrip modification demonstrated successful operation. (33)

On June 25, 1975, Union Carbide Corporation began its evaluation of the PhoStrip process on the full scale plant at Reno/Sparks, Nevada. As with the operation of the pilot plant program, a set of conditions was established on the full scale plant and the system was allowed to come to steady state. After steady operating conditions were achieved, a period of intensive evaluation followed. During this intensive evaluation, Union Carbide maintained a laboratory technician on site to take samples, run analyses, and generally observe the performance of the system. The plant was operated by the staff of the Reno/Sparks Joint Authority.

Table 11 presents a summary of the weekly analyses performed at Reno/Sparks. Because of the nature of the diurnal variation, the analyses were performed on the system during both the high and the low flow periods. The high (8:00 am - 2:00 pm) and the low (2:00 pm - 8:00 am) flow periods are characterized by fairly constant flow rates within each period. This unusual flow variation was a result of the life style and industry in the cities of Reno and Sparks. Figure 17 contains a flow diagram of the system along with the identification letters referred to in Table 11. During this test period, one-third of the plant was operated with the PhoStrip process and the remainder was used as a control. From the schedule in Table 11, it can be concluded that a

TABLE 11. WEEKLY ANALYTICAL SCHEDULE - RENO/SPARKS, NEVADA

LOCATION OR STREAM:		A FEED	B EFFLUENT	C AEROBIC RECYCLE	D ANOXIC RECYCLE	E STRIPPER SUPERNATANT	WASTE SLUDGE	F AERATION BAY 1/3	G AERATION BAY 2/3	H AERATION BAY END	I STRIPPER SLUDGE RECYCLE
Parameter											
Flow	Min. (Avg)	7*	7	7	7	7	7				7
	Max. (Avg)	7	7	7	7	7	7				7
pH	Min.	7G	7G	7G	7G	7G					
	Max.	7G	7G	7G	7G	7G				4G	
TSS	Min.			7+	7+	7+					
	Max.	7C	7C	7+	7+	7+				7	
COD+	Max.	7C	7C								
Temp.	Min.								7		
	Max.								7		
D.O.	Min.	1	1					1	1	1	
	Max.	3	3	7	7	7		3	3	3	
TP+	Min.			7	7,7F	7		3F	3F	3F	
	Max.	7C	7C	7	7,7F	7		3F	3F	3F	
Ortho-P	Min.			7F	7F	7F		3F	3F	3F	
	Max.	7C,F	7C,F	7F	7F	7F		3F	3F	3F	

		SECONDARY CLARIFIER	STRIPPER
Blanket Depth:	Min.	7	7
	Max.	7	7

* = Times/Week
C = Composite
+ = Analyses performed at Tonawanda
G = Grab
F = Filtered
Min = Low flow period of diurnal pattern
Max = High flow period of diurnal pattern

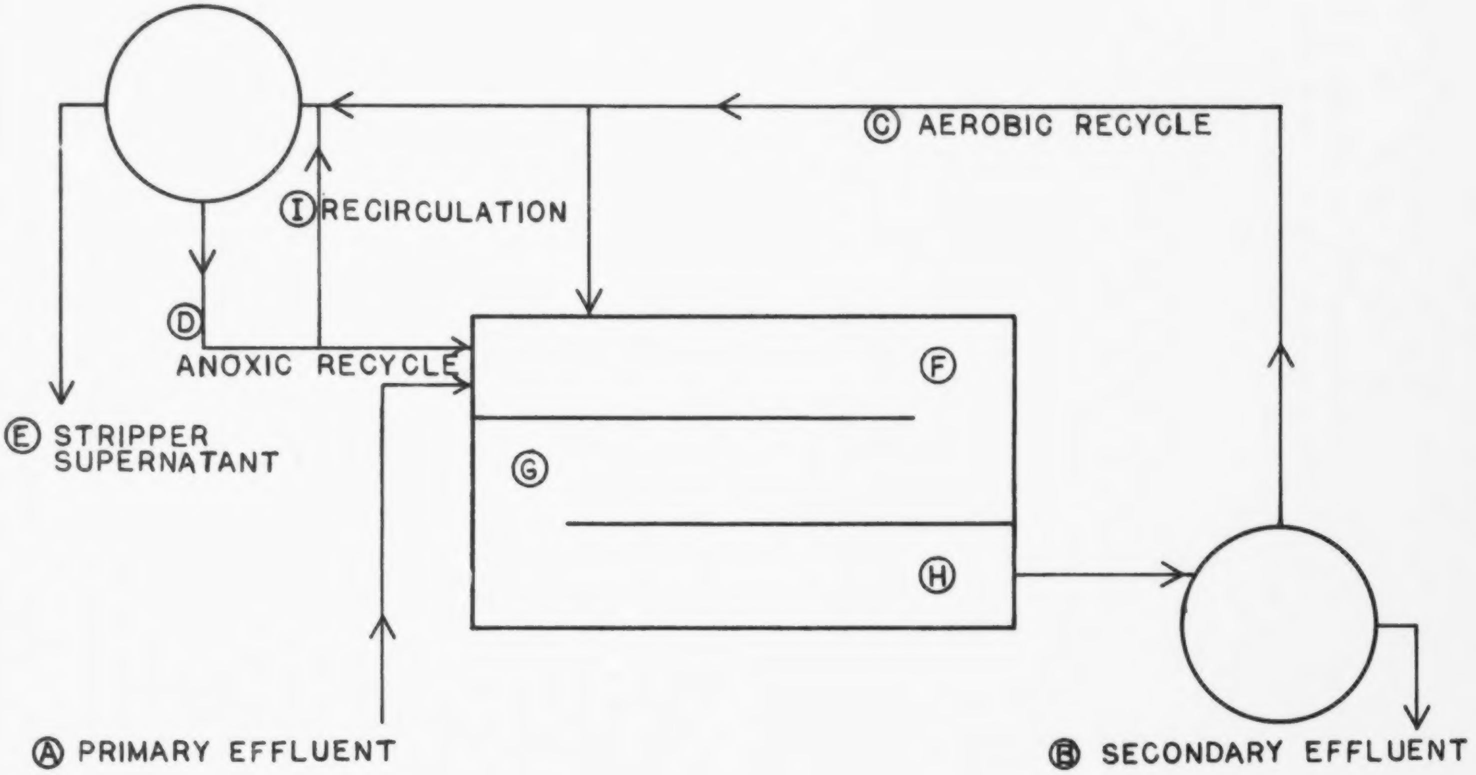


Figure 17. Reno/Sparks - sampling points and flow diagram.

complete evaluation of the system with respect to flows, pH, D.O., TSS, and phosphorus concentration was to be made. An overall evaluation of COD removal was also done.

After becoming familiar with the performance characteristics of the full scale plant, Phase I of the full scale evaluation was begun. The abnormal flow variation (near maximum flow for 18 hrs/day and near minimum flow for 6 hrs each day), coupled with the inability to maintain a blanket level in the secondary clarifier, resulted in less than ideal operating conditions. Conditions similar to those presented in Table 12 were established on August 14, 1975. The relative areas of the secondary clarifier and stripper tank made the maintenance of a sludge blanket in the secondary clarifier impossible. Ideally, the stripper feed and underflow rates should be maintained constant in order to minimize operator control. However, in situations of significant long term flow differences, this can only be accomplished if excess sludge storage capacity is available either in the secondary clarifier or the stripper tank. If the system is maintained with zero blanket level in the secondary clarifier during the high flow period, then, at constant stripper flow rates, the blanket level in the stripper will drop during the low flow period due to an increase in mixed liquor suspended solids. If the blanket level drops such that a minimum required anoxic period is not maintained, the ability of the system to remove phosphorus will be impaired.

There are two ways of maintaining sufficient stripper blanket levels at constant stripper flow rates in systems with prolonged periods of diurnal variation and while maintaining essentially a zero blanket level in the secondary clarifier. The most obvious is to design excess storage capacity into the stripper tank. For example, in the Reno/Sparks situation, a design of 10 hrs anoxic period may be extended to 14 hrs to insure against too short of an anoxic period during low flow conditions. The second technique utilizes the direct recycle stream (R_D in Figure 11) to maintain a constant MLSS level in the aeration tank.

The first control method, excess stripper tankage, represents an increased capital cost which should not be necessary in any but the smallest of plants where operational efforts are generally minimized. The second alternative was not employed at the Reno/Sparks plant because of operational limitations. The capacity of the recycle pumps dictated that higher recycle rates (MLSS levels) were necessary to utilize this scheme. These MLSS levels were not allowable because the characteristics of the wastewater are such that extensive foaming results when the MLSS and associated aeration rates are increased. This situation was observed for the system with and without the PhoStrip modification. The control system used in this case, as well as during the LPE system testing, employed changes in the stripper throughput to maintain a fairly constant MLSS level in the aeration basin and thus maintain a minimum anoxic period in the stripper tank. This situation

TABLE 12. OPERATING CONDITIONS FOR THE PHOSTRIP SLUDGE RECYCLE SYSTEM - FULL SCALE TESTING AT RENO/SPARKS, NEVADA (9/13/75-9/27/75)

Parameter	Minimum	Maximum	Time Weighted Mean
Flow Rates (m^3/day)			
Feed, (Q)	14,690	25,060	22,460
Recycle to Stripper (R_1)	5,180	7,340	6,830
Stripper Underflow (R_3)	2,330	4,150	3,630
Stripper Supernatant	2,850	3,200	3,110
Recirculation Rate (R_2)	4,230	2,420	2,850
Waste Sludge Rate	639	639	639
Aeration Time, Hr. (based on Q)	10.2	6.0	7.1
Anoxic Period, Hr. (based on R_3)	16.	9.4	11.0
MLSS, mg/l	-	-	1,150
MLVSS, mg/l	-	-	900
R_1SS, mg/l	3,060	4,620	4,230
R_1VSS, mg/l	2,480	3,740	3,420
R_3SS, mg/l	9,620	9,120	9,250
R_3VSS, mg/l	7,700	7,300	7,340
TSS Stripper Supernatant, mg/l	65	90	84
VSS Stripper Supernatant, mg/l	50	69	64
D.O., End of Aeration, mg/l	-	1.5 (0.7-2.2)	-
D.O., R_1, mg/l	0.4	0.7	0.6
D.O., R_3, mg/l	0.3	0.3	0.3
pH, End of Aeration Basin	-	7.0	-
pH, R_1	6.8	6.8	-
pH, R_3	6.4	6.4	-
pH, Stripper Supernatant	6.6	6.5	-
Temperature, °C	23	23	23
F/M_a (kg COD/day/kg MLVSS)	-	-	0.96
Clarifier Overflow Rate, m/day	22.4	38.2	34.3
Clarifier Blanket Level, m.	0.08	0.09	0.09

typifies the importance of utilizing an overall systems approach to designing the activated sludge-PhoStrip system. Although the processes will perform well if designed independently, some of the flexibility and ease of operation may be sacrificed. The mode of operation at Reno/Sparks was dictated by the size of the existing tanks, pumps and flowmeters. Had the PhoStrip and activated sludge systems been designed to complement each other, the need to change flow rates around the stripper tank could have been eliminated.

The information presented in Table 12 is representative of the train modified for the PhoStrip process and is consistent with the performance of the control system which was operated independently (appendix). The significant difference in the anoxic retention times for the minimum and maximum flow periods is a result of the change in flow rate of the stream leaving the stripper tank. The volume of the sludge in the stripper tank was maintained at a relatively constant level.

Table 13 presents the average data obtained with respect to COD, suspended solids and total phosphorus. The system demonstrated excellent effluent quality in all respects. The effluent suspended solids level of 13 mg/l compares with the 14 mg/l of the control system. It is interesting to note that the difference between the effluent total and ortho phosphorus coincides closely with that predicted from the product of effluent volatile suspended solids and the P/VSS level for the flow weighted mean. This was mainly due to the fact that the composite samples were made up of periodic grab aliquots which were refrigerated immediately after being taken. The refrigeration significantly reduced the amount of phosphorus released during storage.

The extremely low filtered aeration effluent ortho phosphorus (0.04 mg/l) has been observed in other locations and is an indication of the potential removal capability of the process. The difference between this value and the ortho phosphorus in the final effluent is due to two factors. The filtered aeration effluent is a grab sample while the effluent ortho is a composite. Further, the low D.O. at the end of the aeration basin results in the initiation of phosphorus release in the secondary clarifier. The ortho phosphorus in the effluent is most likely resolubilized polyphosphates. Therefore, lower effluent levels could have been achieved by operating the tail end of the aeration basin at a higher dissolved oxygen level. The effluent quality could have also been improved by increasing the elutriation efficiency. The elutriation efficiency achieved during this period of operation was approximately 38%. Thus, most of the phosphorus released from the sludge in the clarifier was returned to the aeration basin to be taken up again. Higher elutriation efficiencies would have reduced the P/VSS level of the solids in the effluent. As a result, the effluent total phosphorus level would approach the measured ortho phosphorus concentration.

Figures 18 and 19 present a graphical interpretation of the influent and

TABLE 13. ANALYTICAL RESULTS FOR THE PHOSTRIP SLUDGE RECYCLE SYSTEM FULL SCALE TESTING AT RENO/SPARKS, NEVADA (9/13/75-9/27/75)

Parameter	Minimum	Maximum	Flow Weighted Mean
COD, mg/l			
Influent	187	266	253
Effluent	40	37	37
Total Suspended Solids, mg/l			
Influent	-	-	113
Effluent	19	13	13
Effluent Volatile Suspended Solids, mg/l	13	10	10
Total Phosphorus, mg P/l			
Influent	9.6	9.1	9.2
Effluent (ortho)	1.0 (.3)	0.8 (.4)	0.8 (.4)
Stripper Supernatant	44	41	42
Stripper Underflow	257	240	247
Filtered Stripper Underflow	67	55	58
Filtered Aeration Effluent (ortho)	-	- (.04)	-
Secondary Clarifier Underflow	104	161	113
P/VSS			
Secondary Clarifier Underflow	0.042	0.043	0.043
Stripper Underflow	0.025	0.025	0.025

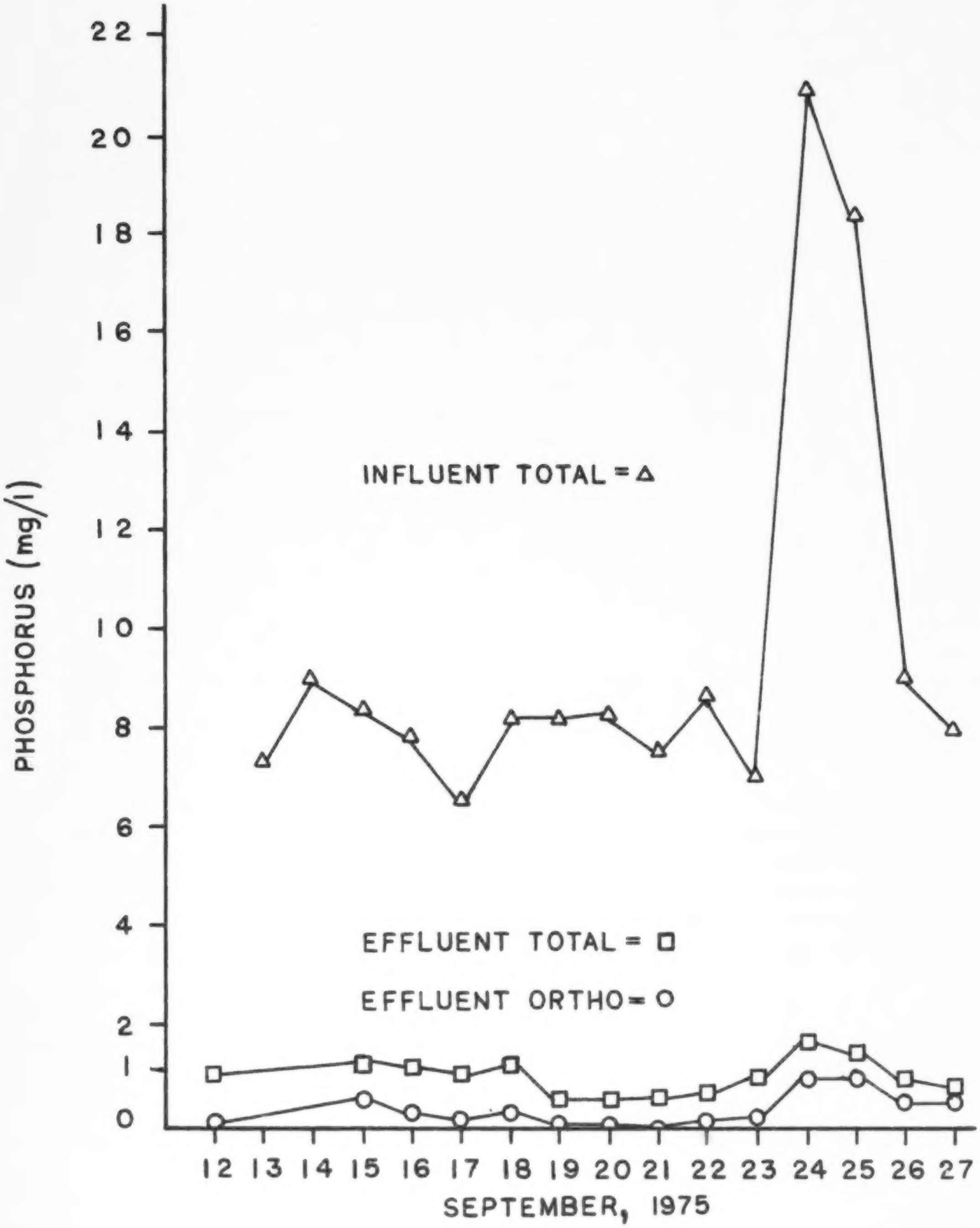


Figure 18. Reno/Sparks full scale phosphorus data obtained during minimum flow period (16% of the total volume treated by the PhoStrip system).

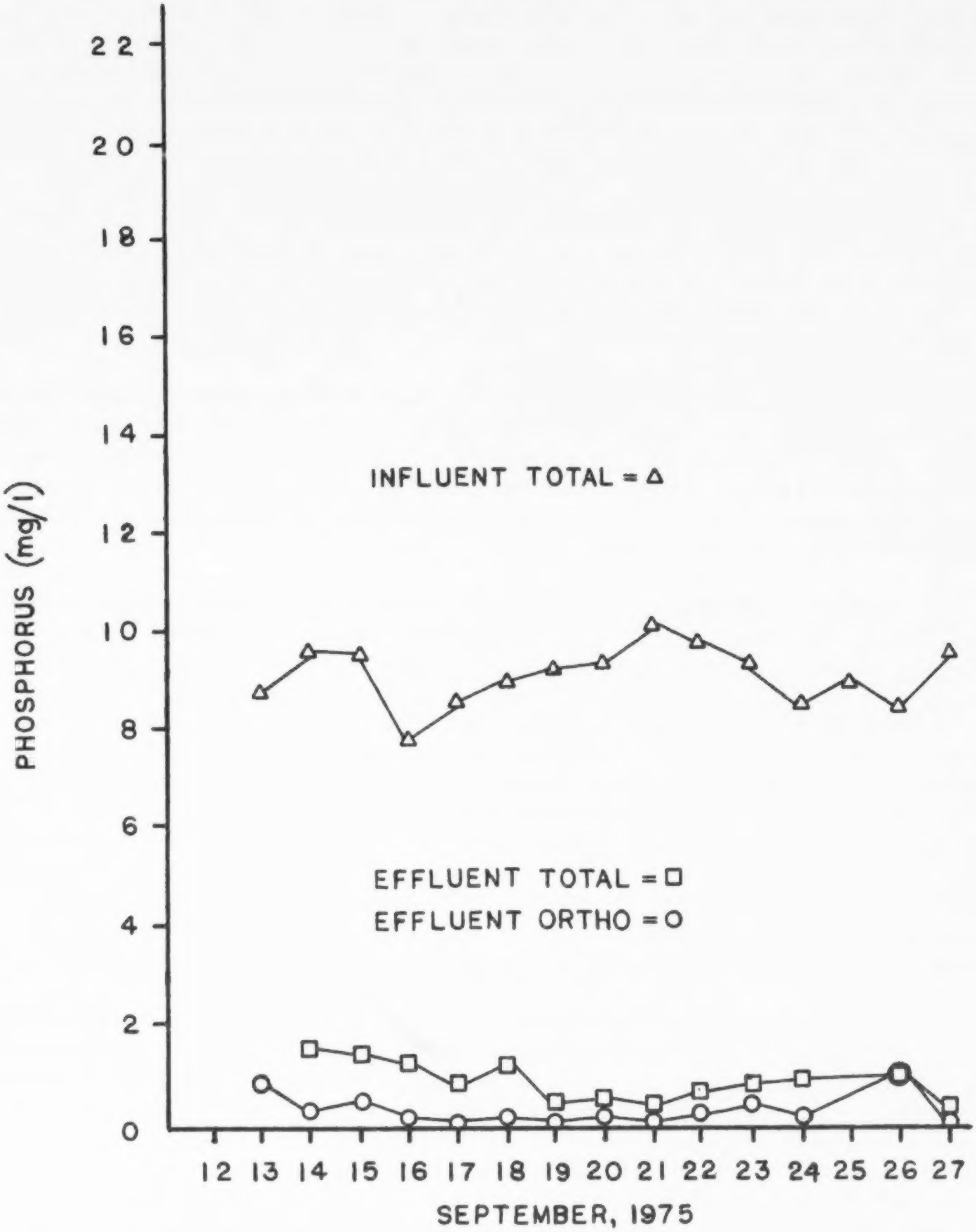


Figure 19. Reno/Sparks full scale phosphorus data obtained during maximum flow period (84% of the total volume treated by the PhoStrip system).

effluent phosphorus data for the minimum and maximum flow periods respectively. The higher effluent total phosphorus data during the first four days was a result of the lower supernatant flow rates employed in the earlier part of the test. Of significance is the fact that the system demonstrated the ability to handle prolonged peak loadings (6 hrs. at 21 mg P/l) without significantly affecting the effluent quality. This is a result of the organisms' ability to take up large quantities of phosphorus over and above the .043 P/VSS level exhibited as an average. The effect of the high influent phosphorus levels observed for September 24 and 25 during the low flow periods did not substantially affect the effluent levels during the maximum flow interval. Since the maximum flow period represents an average of 84% of the liquid treated by the process, the flow weighted effluent phosphorus for these two days is very near 1 mg/l.

The overall phosphorus balance for Phase I of operation is presented in Table 14. Ninety-one percent of the phosphorus was removed from the wastewater. About 63% of the influent was removed through the supernatant stream, while almost 35% was removed with the waste activated sludge. The error in the phosphorus balance is mainly a result of the use of grab samples for the sludge wasting and accumulation term.

A comparison between the control system and the PhoStrip system at Reno/Sparks has been published by Peirano (31) at various operating conditions. Therefore, extensive testing of the control system was not performed during this evaluation. However, Table 15 presents data used to determine the P/VSS level in the sludge from the control system during the period of operation under discussion. The calculations presented in the table show that phosphorus content of the sludge in the control was approximately 2.4% of the VSS. Since approximately the same mass of volatile suspended solids was wasted from the PhoStrip and control units, and since the phosphorus loading (kg/day) to the two units was identical, the control system removed approximately 19.5% of the influent phosphorus. This removal was accomplished through sludge wasting alone. This is equivalent to an effluent phosphorus level of 7.4 mg P/l for the control system during this phase of operation.

A comparison of the data between the full and pilot scale systems run in the sludge recycle mode indicates a significant improvement in performance for the full scale system. At similar operating conditions the full scale system achieved an effluent phosphorus level less than 1.0 mg/l while the pilot system was able to achieve only 3.5 mg P/l of effluent. The phosphorus balances for the two systems indicate that both systems removed nearly the same fraction of the influent phosphorus through wasting of the activated sludge. However, the pilot unit removed only 20.9% of the influent phosphorus through the supernatant while the full scale system removed 63.2% of the phosphorus through this stream. This difference is mainly a result of the performance of thestripper tanks with respect to elutriation efficiency. In the

TABLE 14. PHOSPHORUS BALANCE DATA FOR THE FULL SCALE TESTING AT RENO/SPARKS, NEVADA PHOSTRIP SLUDGE RECYCLE SYSTEM

Influent Phosphorus, kg/day	206.7
Effluent Phosphorus, kg/day	18.0
, % of Influent	8.7
Stripper Supernatant P, kg/day	130.6
, % of Influent	63.2
Waste Plus Accumulated P, kg/day	72.2
, % of Influent	34.9
Phosphorus Balance Error, % of Influent	+6.8

TABLE 15. SLUDGE PHOSPHORUS CONCENTRATION OF THE CONTROL SYSTEM - FULL SCALE TESTING AT RENO/SPARKS, NEVADA

Date	TSS, MG/L		VSS, MG/L		TPO_4-P, MG/L	
	ML	Recycle	ML	Recycle	ML	Recycle
9-15-75	1452	6260	1060	4744	35	116
9-17-75	1276	5220	1022	4290	37	110
9-19-75	1044	6084	844	5676	22	142
Mean	1260	5850	980	4900	31	122

$$\% \text{ P in solids } = \frac{[(TP) - (SOL\ TP)]\ 100}{VSS}$$

Mixed Liquor SOL TP $\approx$ 7.2 mg/l*

$$\text{Mixed Liquor } = \frac{(31.3 - 7.2)\ 100}{980} = 2.4\%$$

$$\text{Recycle } = \frac{(122 - 7.2)\ 100}{4900} = 2.3\%$$

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* obtained from a mass balance around control system.

pilot system only 15% of the phosphorus which was released by the sludge in the stripper made its way to the supernatant. Thirty-eight percent of the phosphorus released in the full scale stripper was removed through the supernatant stream. Since the two stripper tanks were run at nearly the same conditions, the improved elutriation efficiency was a result of the change in scale. The sludge recycle system runs more efficiently at higher tank diameter-to-depth ratios. The higher ratios, coupled with the design of the stripper influent mechanism used at Reno/Sparks, may have caused the solids in the stripper tank to roll somewhat. As a result, the released phosphorus sludge was more evenly distributed throughout the tank and thus could more easily diffuse to the supernatant. The theoretical elutriation efficiency for the conditions of the full scale system would predict an $\epsilon = 0.22$ according to Figure 9. For the purpose of designing, it is recommended that the theoretical elutriation efficiency curves be used because of the uncertainty of the effect the sludge distribution device had on the performance of the stripper tank.

FULL SCALE TESTING OF THE LPE SYSTEM

The evaluation of the full scale LPE system was originally to last twelve months and was to be divided into four phases. The first phase was preliminary testing of 50,000 m^3/day. The second was a long term intensive evaluation at 50,000 m^3/day. The third and fourth phases were to test performance at 76,000 m^3/day. As was discussed earlier, the duration of the program was cut due to a water quality study on the Truckee River. As a result, only the preliminary 50,000 m^3/d program was completed.

During PhoStrip operation the raw degritted sewage was settled in the number 2 and 3 primary basins and then evenly split and discharged to the three aeration basins. Each basin is split into three tanks and operated in a serpentine plug flow manner. After aeration, the mixed liquor was discharged to three independent secondary settling tanks. The effluent was post-aerated and chlorinated prior to discharge.

The PhoStrip system was prepared for the demonstration project during August and September of 1976. An elutriation pump was installed so that primary effluent could be used for washing the solubilized phosphorus from the sludge in the stripper tank. An additional pump was purchased and installed for the purpose of removing the phosphorus-rich supernatant from the stripper and discharging it to the head end of the third aeration system. The number 1 and 2 systems operated on PhoStrip while the number 3 did not.

The return activated sludge from the number 1 and 2 secondary clarifiers was pumped to the stripper with some sludge bypassed directly to the aeration basins. Sludge from the stripper was pumped to and split evenly betwen the number 1 and 2 aeration basins, thus completing the cycle.

Figs. 20 & 21 diagramatically present the configuration of the PhoStrip system during this phase of the demonstration project. All flows were metered and volumes integrated. Suitable valves were installed for adjusting flows to and from the stripper and in the secondary sludge return bypasses to the aeration basins.

Sludge depths in the secondary clarifiers and the anoxic stripper were measured six times daily and used to control stripper sludge flow and wasting rates. It was important to maintain adequate sludge blankets in the secondary clarifiers so that sludge would properly thicken. The stripper was operated to maximize anoxic retention volumes (less than 0.9 m freeboard water).

The initial period of operation was for the purpose of optimizing overall system performance, minimizing the volume of supernatant and acclimating to cold weather conditions. Samples were taken and analyses performed in the Reno/Sparks treatment plant laboratory and at Union Carbide's Tonawanda laboratory. The analytical schedule is included as Table 16. Additional data were taken as a normal part of the plant's monitoring schedule. Flow rates in the PhoStrip system were noted daily.

A Union Carbide engineer was located at the facility during this project. His responsibilities were to supervise the operation of the retrofitted PhoStrip system, collect samples and perform necessary analyses.

The PhoStrip system had been initially started up during October 1976 and operated by plant personnel until the start of the project on November 22, 1976. From November 22, 1976 to early January 1977 various operating conditions were tested to determine the response of the system to different modes of operation. On January 13, 1977 the conditions were changed to those necessary to achieve performance for the planned Phase II intensive evaluation. The data presented in Tables 17, 18 and 19 are from the steady performance achieved during Phase I prior to shutting down the system.

Table 17 presents the steady state operating conditions obtained prior to the termination of the program. As with the sludge recycle case, the fact that the PhoStrip system was a temporary add-on resulted in operations that were not optimum. Again, the flow rates around the stripper tank had to be changed as a function of the high and low flow periods during the day. This flow rate change was necessitated by a change from the influent flow patterns normally observed at the plant. Normally, the speeds of the two pumps in the wet well of the pump station that feeds the plant are automatically varied to pace with the incoming flow. However, during the demonstration project the motor speed control centers were inoperative and the pumps could only be operated at full speed. By using the interceptors as a continuation of the wet well, the plant was operated in a one pump/two pump mode. This method of operation resulted in step changes in the flow rates between the minimum and

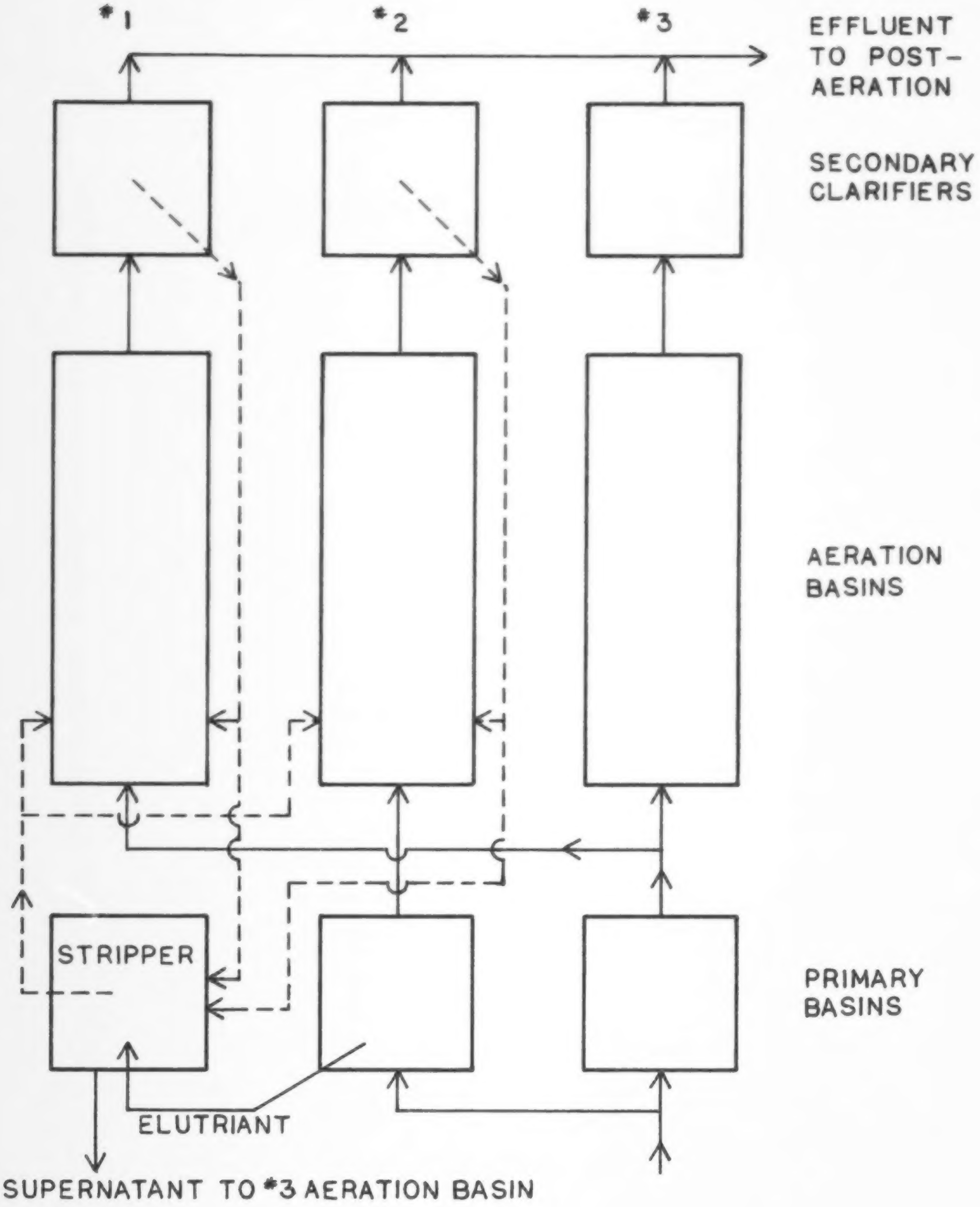

Figure 20. Flow diagram of PhoStrip LPE system at Reno/Sparks, Nevada.

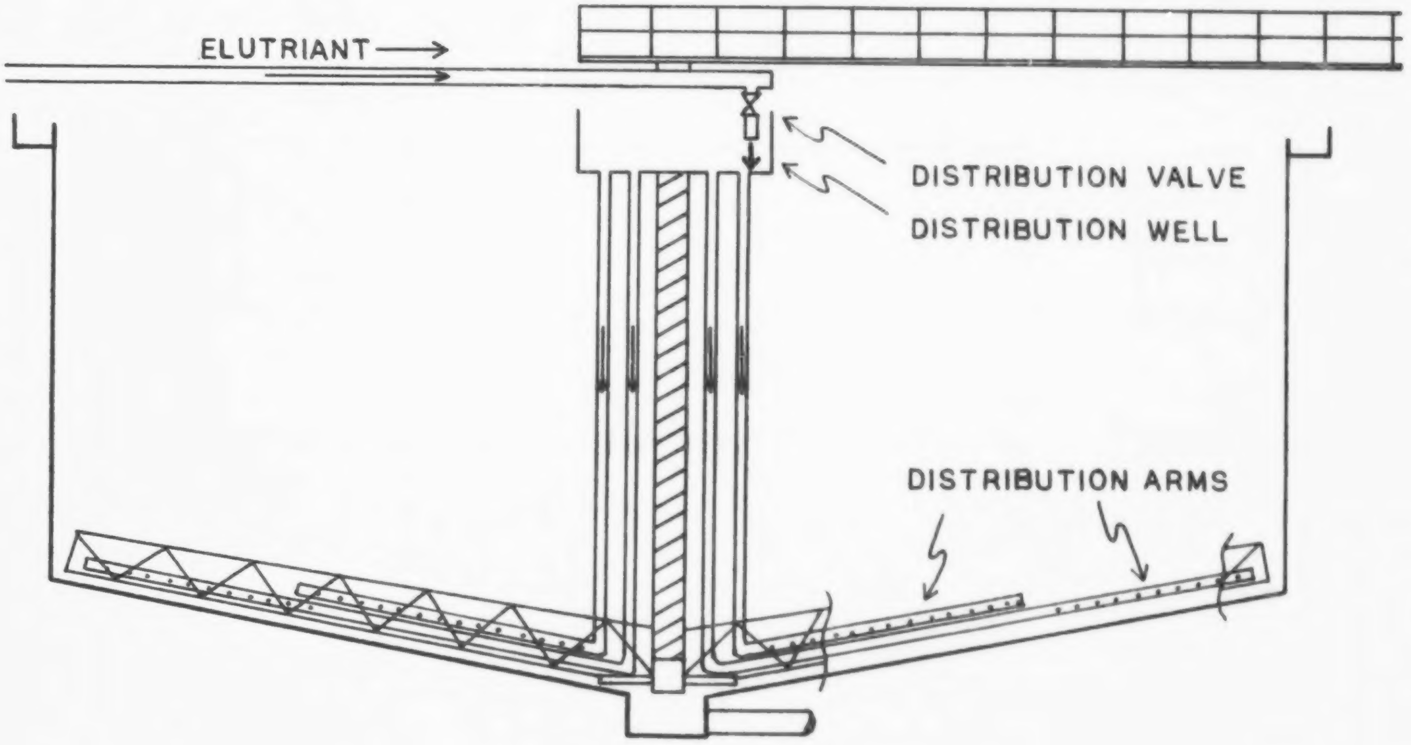

Figure 21. Stripper tank and elutriant distribution system.

TABLE 16. ANALYTICAL SCHEDULE FOR FULL SCALE TESTING AT RENO/SPARKS, NEVADA - PHOSTRIP LPE SYSTEM

SAMPLE LOCATION	BOD_5	TSS	VSS	pH	TKN	TP	OP**	TYPE
Primary Effluent	2+	7	1	5		5	5	C
Secondary Effluent								
Clarifier No. 1	2	7	1	5		5	5	C
Clarifier No. 2	2	7	1	5		5	5	C
Aeration Basin (end)*								
Tank No. 1		7	1	3			3	G
Tank No. 2		7	1	3			3	G
Stripper Supernatant	1	2	2	5		5	5	C
Anoxic Sludge Recycle*		1	1			1	3	G
Aerobic Sludge Recycle*								
Tank No. 1		7	1			1		G
Tank No. 2		7	1			1		G

+ = analyses/week
C = 24 hr. composite
G = grab
* = two grab samples: one at high flow, one at low flow
** = mixed liquor soluble ortho phosphates were measured 3 times per week

(Flow rates are normally taken as part of plant operating data.)

TABLE 17. OPERATING CONDITIONS FOR FULL SCALE TESTING
AT RENO/SPARKS, NEVADA
PHOSTRIP LPE SYSTEM (JANUARY 18-30, 1977)

PARAMETER	MINIMUM TRAIN 1/TRAIN 2	MAXIMUM* TRAIN 1/TRAIN 2	FLOW WEIGHTED MEAN TRAIN 1/TRAIN 2
Flow Rates, m^3/day			
Feed Q, (each train)	17280/17280	29380/29380	22460/22460
Recycle to Stripper, R_1	1987/2160	3200/3370	2510/2680
Stripper Underflow, R_3	1210/1470	2250/2250	1640/1810
Stripper Supernatant, S	6570	7000	6740
Elutriation, EL	5270	5620	5440
Total Aerobic Recycle, $R_1 + R_{direct}$	-	-	4490/5530
Sludge Wasting	-	-	331/418
Aeration Time, Hrs. (based on Q)	9.0/9.0	5.1/5.1	6.8/6.8
Anoxic Period, Hrs. (based on R_3)	11.6	7.0	9.5
MLSS, mg/l	1230/1280	1110/1140	1170/1220
VSS/TSS	0.8	0.8	0.8
R_1SS, mg/l	5750/5460	6770/6720	6210/6030
VSS/TSS	0.78	0.78	0.78
R_3SS, mg/l	8930	8310	8650
R_3VSS, mg/l	7360	6590	7010
TSS Stripper Supernatant, mg/l	-	-	59
VSS Stripper Supernatant, mg/l	-	-	54
pH, Mixed Liquor, mg/l	7.1/7.0	7.0/7.0	-
Temperature (Influent), °C	13	13	13
F/M_a (kg BOD_5/day/kg MLVSS)	0.45	0.88	0.62
Secondary Clarifier Overflow Rate, m/day	26.4/26.4	44.8/44.8	34.3/34.3
Clarifier Blanket Level, m.	-	-	< 0.2

* duration at max. flow $\approx$ 11 hrs/day

maximum flow period as well as an increase in the duration of the low flow period from ~6.0 to ~13 hours.

With the LPE mode of operation, it is desirable to operate without any thickening occurring in the stripper tank. A comparison of elutriation flow rate and the stripper supernatant flow indicates that this was not accomplished during this phase of the program. The clarifier and stripper underflow suspended solids levels further confirm this point. Attempts to increase the blanket level in the secondary clarifier above 0.2 m were very successful with respect to increasing the R_1SS level to that of the R_3SS concentration. However, the RSR (rapid sludge return) units designed into the secondary clarifier were not able to handle the higher TSS levels. The arms of the units would plug and the return sludge flows were disrupted.

The retention time in the stripper was limited by the amount of volatile solids that had to be put through the anoxic period to achieve the desired phosphorus removal and the size of the primary clarifier employed as the stripper tank. Slightly longer (increased 1 to 2 hr) would be more desirable for the operating temperature of this system (13°C).

The average analytical results observed during this period of operation are contained in Table 18. The effluent BOD_5 and suspended solids levels are typical of those obtained from activated sludge systems and are comparable with those obtained from reactor #3 which was not operating in the PhoStrip mode (Appendix). The phosphorus data indicates that the LPE system was able to achieve a high quality effluent while treating two-thirds of the total plant flow. As with the pilot plant data, the difference between the effluent total and ortho phosphorus levels does not agree with that predicted by the effluent total and ortho phosphorus levels and does not agree with that predicted by the effluent volatile suspended solids and the P/VSS ratio of the clarifier underflow. Because of the larger number of samples taken with two trains operating in the PhoStrip mode, it was not possible to refrigerate the effluent samples. Consequently, a large fraction of the ortho phosphorus in the effluent samples resulted from resolubilization of the polyphosphate within the effluent suspended solids. The other factors mentioned earlier may have also contributed to this difference.

Figure 22 graphically displays the influent and effluent phosphorus levels achieved during the two week steady state period. This effluent quality could have been improved by running the system at longer stripper SRT levels. This would have resulted in more phosphorus release in the stripper tank and, therefore, more phosphorus removed with the supernatant. Increasing phosphorus removal through the supernatant reduces the P/VSS level of the sludge and would result in less effluent phosphorus due to suspended solids. The size of the stripper tank at Reno/Sparks did not permit operations at longer retention times with the required sludge throughput. Reducing the sludge

TABLE 18. ANALYTICAL RESULTS FOR FULL SCALE TESTING AT RENO/SPARKS, NEVADA PHOSTRIP LPE SYSTEM (JANUARY 18-30, 1977)

PARAMETER	MINIMUM TRAIN 1/TRAIN 2	MAXIMUM TRAIN 1/TRAIN 2	FLOW WEIGHTED MEAN TRAIN 1/TRAIN 2
BOD_5, mg/l			
Influent	-	-	168
Effluent	-	-	28/29
Total Suspended Solids, mg/l			
Influent	-	-	102
Effluent	-	-	21/23
Effluent Volatile Suspended Solids, mg/l	-	-	~18
Total Phosphorus, mg P/l			
Influent	-	-	9.1
Effluent (ortho)	-	-	0.8(.5)/0.9(.6)
Stripper Supernatant	-	-	36
Stripper Underflow	255	270	260
Filtered Stripper Underflow	50	70	59
Secondary Clarifier Underflow	215/195	280/250	245/220
P/VSS			
Secondary Clarifier Underflow	.048/.046	.053/.048	.05/.047
Stripper Underflow	0.028	0.030	0.029

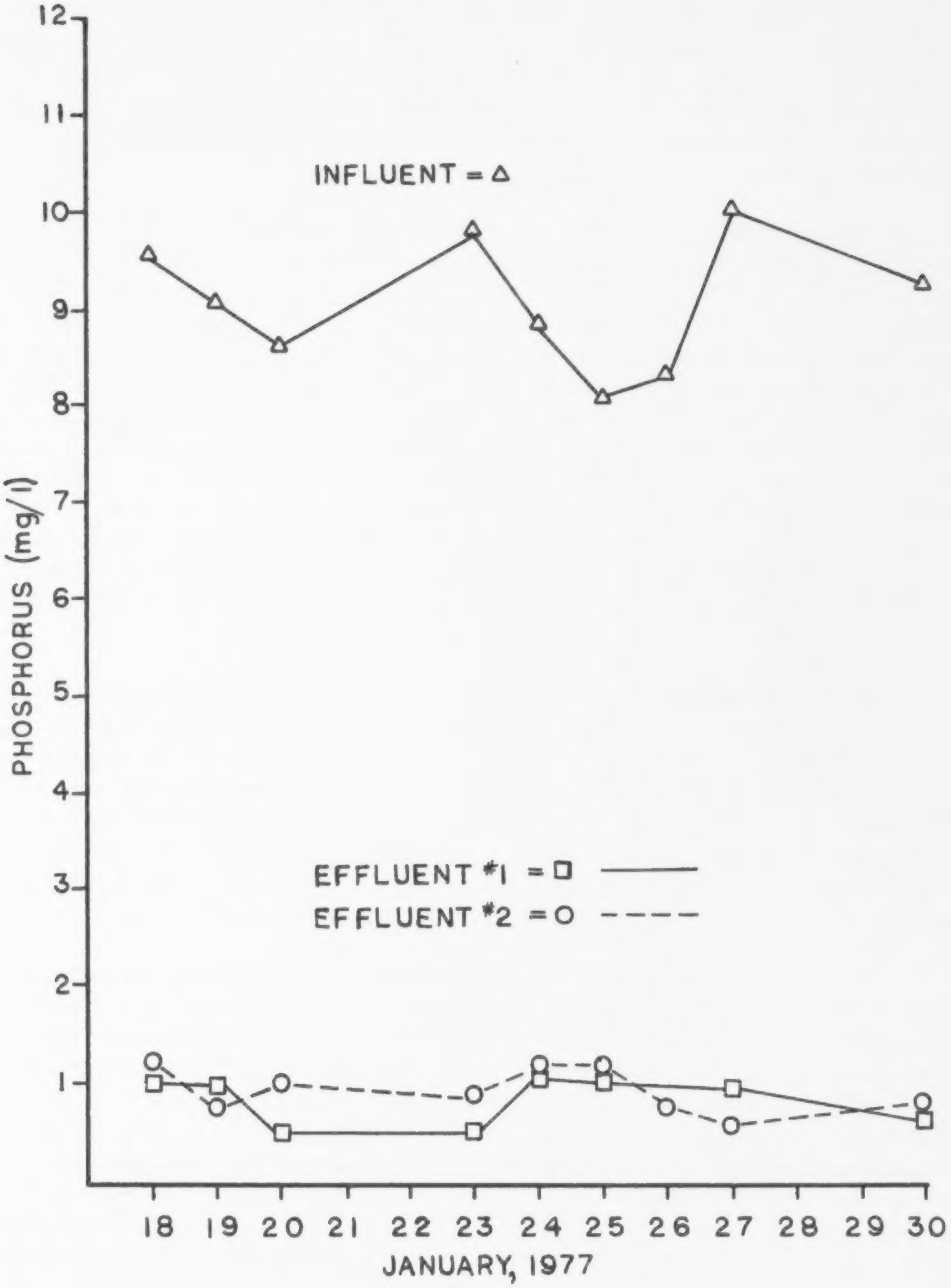

Figure 22. Reno/Sparks full scale phosphorus data for LPE system.

throughput would increase the retention time but would also increase the amount of phosphorus that would have to be released per unit of mass sludge transferred to the stripper. Thus, only a modest reduction, if any, in the P/VSS level could have been achieved in this manner. A stripper tank with the same cross sectional area but larger volume (more depth) would have been a more effective system.

The phosphorus removed through the supernatant could have been increased by improving the elutriation efficiency of the system. As the system was operated, nearly 59% of the phosphorus released in the stripper tank appeared in the stripper supernatant. From Figure 9 it can be seen that an increase in elutriation efficiency can be achieved by increasing the elutriation flow rate. An increase in elutriation efficiency done in this manner results in an increased chemical cost due to a larger stripper supernatant flow. The results of this phase of operation represent the minimum elutriation flow allowable for the tankage volume and elutriation distribution system used. A taller stripper tank would have allowed the elutriation flow rate to be reduced. A stripper tank with a smaller cross sectional area or a sludge return system that could handle higher solids concentration would have resulted in less thickening and, therefore, a further reduction in the supernatant flow rate and the associated chemical costs.

Table 19 summarizes the results of the mass balance performed on the data obtained during this phase of operation. Unlike the data for the sludge recycle system, the influent phosphorus mass rate depends on the feed to the aeration basin and the elutriation flow rate because primary effluent was used as the elutriant. The high P/VSS of the system manifests itself in the amount of phosphorus removed through activated sludge wasting. As indicated earlier, most of the error in the mass balance is in the waste plus accumulation term because of the necessity for using grab samples on these streams.

A comparison between the LPE and sludge recycle systems indicates that the LPE system was capable of producing the same quality effluent while utilizing the same stripper tank, but handling twice the flow at a lower temperature. The main reason for this is an improvement in elutriation efficiency with the LPE system ($\varepsilon = 0.59$ vs. $\varepsilon = 0.38$). This increase in elutriation efficiency was achieved with no significant increase in the percent of influent flow appearing in the supernatant from the stripper. The LPE system removed more phosphorus through activated sludge wasting than the sludge recycle system due to the higher P/VSS ratio in the sludge. However, the P/VSS ratio is a controllable parameter which was restricted by the conditions resulting from the requirement of using existing tankage.

TABLE 19. PHOSPHORUS BALANCE DATA FOR FULL SCALE TESTING AT RENO/SPARKS, NEVADA PHOSTRIP LPE SYSTEM (JANUARY 18-30, 1977)

Influent Phosphorus*, kg/day	458.3
Effluent Phosphorus, kg/day	38.2
, % of Influent	8.3
Stripper Supernatant P, kg/day	242.6
, % of Influent	52.9
Waste Plus Accumulation P, kg/day	168.6
, % of Influent	36.8
Phosphorus Balance Error, % of Influent	- 2.0

* feed to aeration plus elutriation

PROCESS DESIGN CONSIDERATIONS

Kinetics

As was mentioned in Section 4, there are two major considerations restricting the performance of the PhoStrip process with respect to uptake and release: kinetics, and the phosphate saturation level in the sludge. Once the kinetics of the process are understood, the use of mass balances as one of the design tools can result in a design which prevents the system from approaching too near the saturation level of the sludge. Designing a system for a P/VSS ratio of less than 0.04 will insure ample safety with regard to saturation level and produce an effluent quality of less than one mg phosphorus per liter for most activated sludge systems.

The kinetic rate values for the data obtained from both pilot and full scale operations are presented in Table 20. These values were obtained by assuming zero order reactions relative to the soluble phosphate concentration. (23-25) Data from the preliminary operational periods of the LPE System were also used to help define these values. Figure 23 shows the relationship obtained for the uptake rate values as a function of temperature. It is recommended that this relationship be utilized in the design of activated sludge systems at F/M levels ranging from 0.3 to 0.7 kg BOD_5/day/kg MLVSS.

The data in Table 20 indicates that the full scale LPE system produced a release rate (k_R) that was nearly 100% greater than that of the pilot scale LPE system. A comparison between the pilot and full scale sludge recycle systems indicates that there was no significant difference between the two units with respect to release rate values. Since the full scale sludge recycle study exhibited nearly the same k_R value as the pilot scale LPE system, the full scale LPE system was more efficient than the sludge recycle system with respect to phosphorus release. The exact explanation for this cannot be obtained from these data. It is recommended that a k_R value of .03 be utilized for designing PhoStrip systems with anoxic sludge retention times ranging between 5.5 and 15 hours. This rate represents a significant safety factor for the LPE system but none for the sludge recycle system.

Elutriation Efficiency

As was discussed earlier, the sludge recycle system operating on the full scale system exhibited significantly higher elutriation efficiencies than the pilot scale system and moderately higher than the theoretical model would predict. The use of the theoretical model provides a 1.7 safety factor over the full scale data and is therefore recommended for designing the process.

The efficiency of the LPE system for the full scale demonstration study is compared to the pilot scale correlation in Figure 24. Generally, the pilot

TABLE 20. KINETIC CONSTANTS AND ELUTRIATION EFFICIENCIES AT RENO/SPARKS, NEVADA (NOVEMBER 1976 - JANUARY 1977)

	PILOT PLANT			FULL SCALE								
					LPE WEEK BEGINNING							
KINETIC VALUES	SLUDGE RECYCLE PHASE I	PHASE II	PHASE III	SLUDGE RECYCLE STEADY STATE PHASE	11/21	11/28	12/5	12/12	12/26	1/2	1/9	STEADY STATE PHASE
k_u	.06	.05	.06	.07	I.D.	I.D.	.06	I.D.	.05	.05	.05	.05
k_R	.03	.02	.02	.03	.06	.05	.05	I.D.	.05	.04	.04	.04
ε	.15	.71	.58	.38	.36	.36	.38	I.D.	.68	.55	.49	.59
γ	1.0	1.4	.8	.9	.8	.9	.6	.9	2.7	1.1	.9	1.6
Mixed Liquor Temp., °C	16	15	17	23	15	12	13	12	12	13	13	13
RT_s, Hrs.	8.0	6.0	5.5	11.0	6.6	6.8	5.9	5.9	14.4	12.7	10.6	9.5

k_u = uptake rate kg P/day/kg MLVSS
k_R = release rate kg P/day/kg VSS in stripper
ε = elutriation efficiency
γ = supernatant flow/stripper underflow, or elutriation flow/stripper underflow
RT_s = retention time in stripper based on R_3

Notes:

No data was taken on 12/19.

I.D. = insufficient data

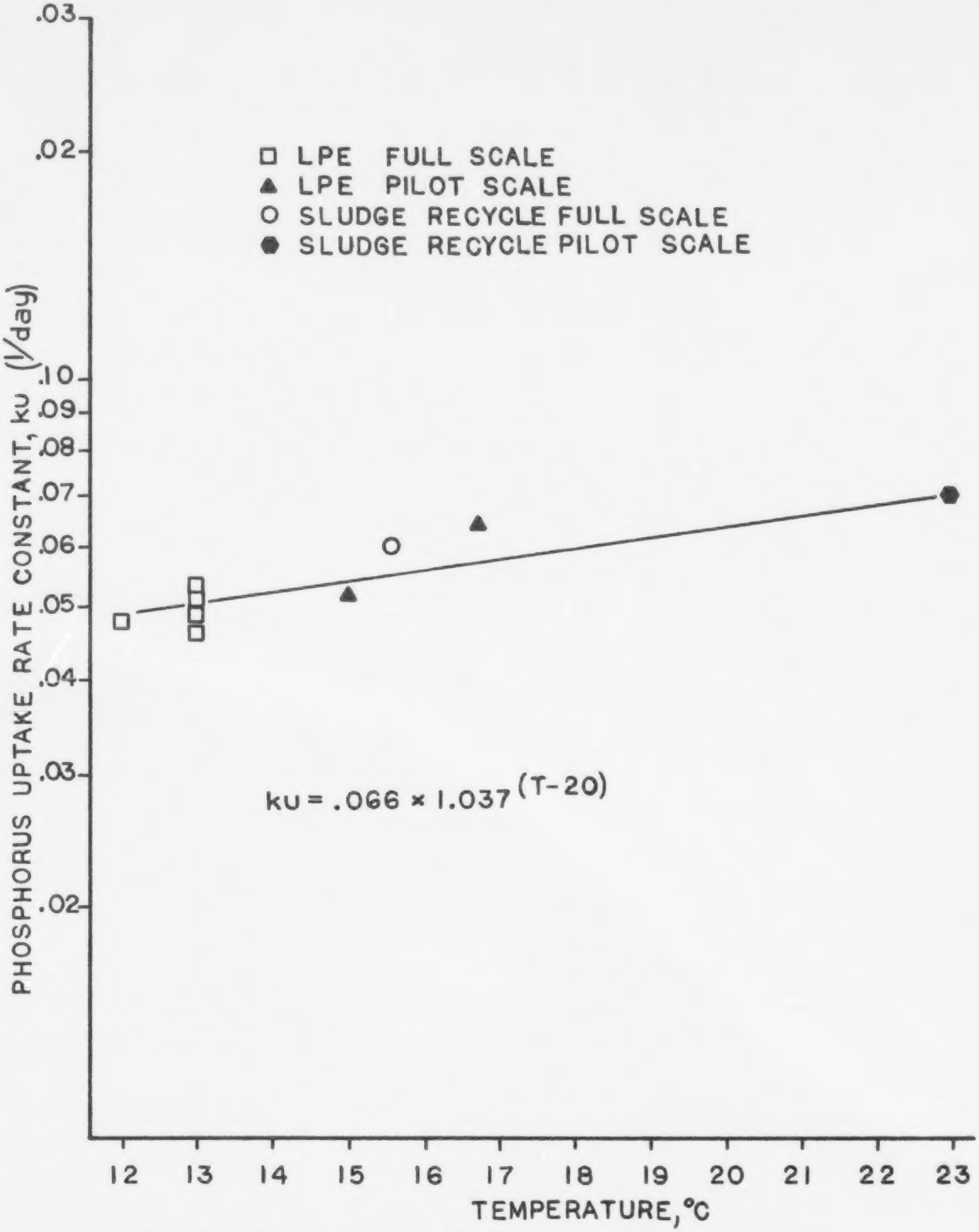

Figure 23. Phosphorus uptake rate constant versus temperature (Reno/Sparks, Nevada).

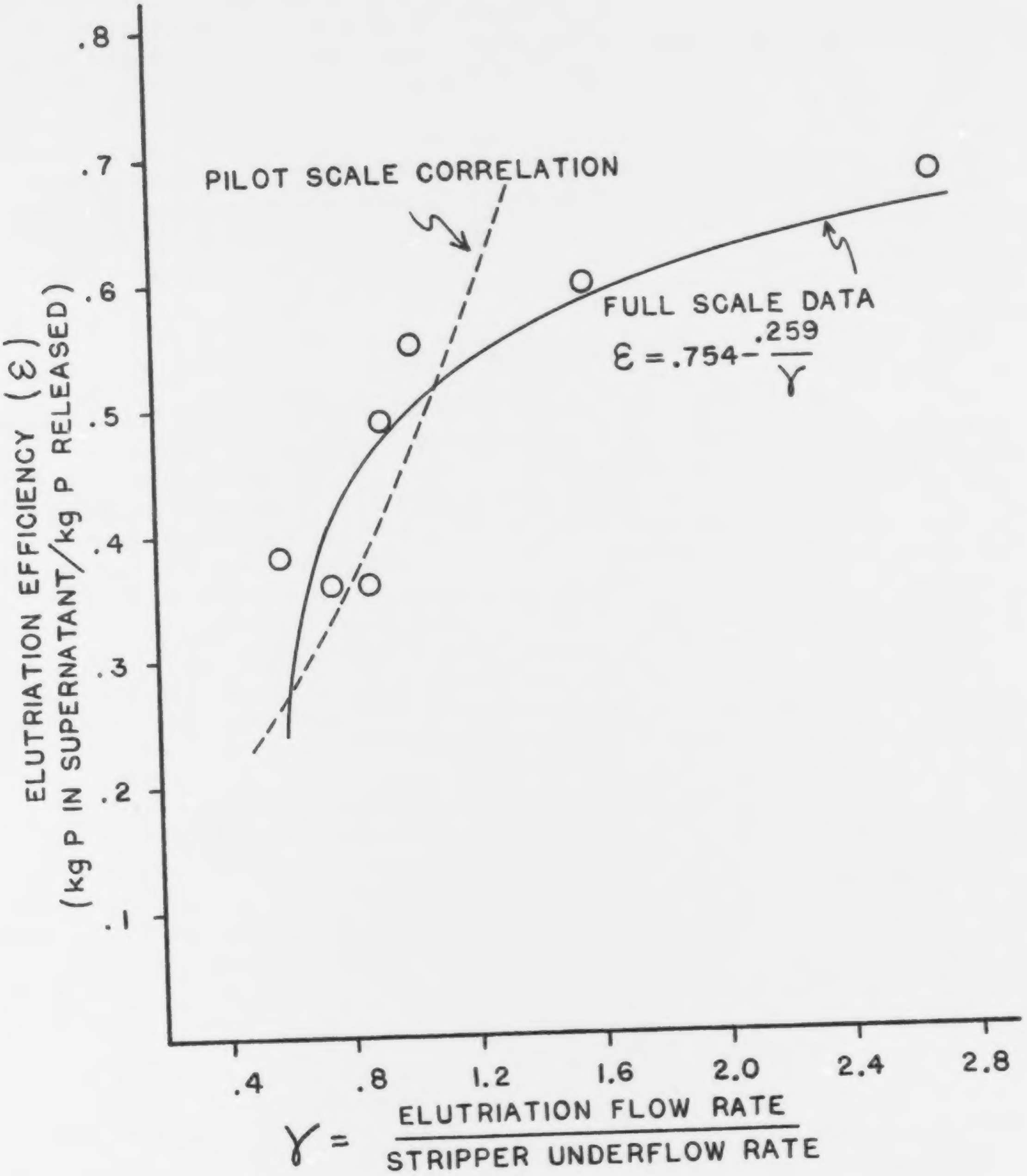

Figure 24. Elutriation efficiency for the LPE modification.

and full scale data compare rather well at low γ values while significant deviation occurs when γ exceeds 1.1. Figure 24 also contains an equation relating elutriation efficiency to γ . This equation was obtained from a least-squares curve fit of the data, and is recommended for design usage. The safety factor for the LPE system is incorporated in the release rate constant and is equivalent to a safety factor of 2.0 for the overall process design. Theoretically, the elutriation efficiency for the LPE system will improve with increased stripper height at the same anoxic sludge residence times (smaller cross section area). It is anticipated that systems with more than the 2.0 m of sludge blanket available during this study will exhibit higher efficiencies than those shown in Figure 24.

Settling Characteristics of the Biological Sludge

In order to adequately design the PhoStrip process, it is necessary to know the settling characteristics of the sludge so that the stripper cross sectional areas can be determined for a desired throughput. Further, the optimum performance of the LPE modification can only be achieved if no increase in suspended solids concentration is allowed to occur from the stripper feed stream to the stripper underflow. Therefore, best performance can be achieved with the minimum possible stripper cross sectional area which allows for sufficient anoxic sludge residence time and mass throughput. If safety factors are applied to stripper tank design for this system, they should result in an increase in stripper height, not cross sectional area. However, the safety factor incorporated into the release rate is sufficient that no safety factor should be applied directly to the sizing of the stripper tank.

Ideally, the settling constants for the design of the stripper tank should be obtained from a PhoStrip pilot plant program. In lieu of such data, it can be assumed that the settling constants are equivalent to those characteristic of the activated sludge system to be employed. Again, the safety factors in the elutriation efficiency for the sludge recycle system and the release rate constant for the LPE system will be sufficient to provide a safe design if reasonable settling constants are employed.

SECTION 7

LIME REQUIREMENTS, CHEMICAL SLUDGE PRODUCTION, AND CHEMICAL SLUDGE DEWATERING

Precipitation of phosphorus utilizing lime (CaO) is an integral part of the PhoStrip process. As discussed in Section 4, there are only a limited number of cases where low effluent phosphorus levels are achievable without the use of a chemical precipitant. The example presented in Figure 6 showed that nearly 82% of the phosphorus removed from the raw wastewater was concentrated in the waste chemical sludge.

The principal forms of phosphorus in wastewaters are the ortho phosphate ion, polyphosphates, and organic phosphates. The activated sludge process converts the organic phosphates and polyphosphates to the ortho form through biochemical activity. In the PhoStrip process, this ortho phosphorus is taken up in the aeration basin and released back into solution in the anoxic stripper tank. Thus, phosphorus in the ortho form is the material that is precipitated from the stripper supernatant stream.

Lime is employed as the precipitant in this process because of its low cost and the fact that phosphate precipitation with lime is pH dependent and independent of the phosphorus concentration. As presented in the EPA Process Design Manual for Phosphorus Removal (34), the relationship of total residual phosphorus with pH for secondary effluents is given by equation (14).

$$\text{Log } P = 3.51 - .392\,(\text{pH}) \tag{14}$$

Since the stripper supernatant stream is high in phosphorus content (30-80 mg P/1) and low in volume (generally 5 to 15% of the wastewater flow), the PhoStrip process requires only a fraction of the lime which would otherwise be required if phosphorus were removed by direct addition of lime to the wastewater.

Because the supernatant stream does have a higher phosphorus content, a greater fraction of the lime added is consumed in the phosphate precipitation reactions than when lime is added to the raw wastewater. The chemical sludge produced has a higher fraction of phosphorus, thus significantly reducing the quantity of sludge produced. Further reductions in sludge production are achieved by maintaining a relatively low pH level for precipitation ($8.5 \leq pH \leq 9.0$). At these pH's, comparatively little calcium carbonate is precipitated and almost no magnesium hydroxide is formed. The sludge produced is largely calcium phosphate precipitates.

The chemistry of phosphorus precipitation is very complex. For the sake of simplicity, it is assumed that calcium ion reacts with the HPO_4^{-2} phosphorus species to form calcium hydroxyapatite according to equation (15). (34)

$$3HPO_4^{-2} + 5Ca^{+2} + 4OH^{-1} \rightarrow Ca_5(OH)(PO_4)_3 + 3H_2O \qquad (15)$$

The solubility of this precipitate is so low that at pH levels as low as 8.5 soluble phosphorus levels of much less than 1 mg/l are achievable. Equation (14) predicts higher phosphorus levels than that obtained from the solubility of calcium hydroxyapatite due to the inaccuracies of assuming that the calcium phosphate species in equation (15) is the only product of phosphate precipitation. It is likely that both octacalcium phosphate and tricalcium phosphate are formed preferentially as the initial precipitates. (35) This is possible, even though these are more soluble products, because the kinetics of formation of these materials is much faster than that of the formation of calcium hydroxyapatite. However, as time progresses after the initial precipitate is formed, these more soluble species are converted to the hydroxyapatite form. (36) Therefore, it can be assumed that the hydroxyapatite form is the final end product of calcium phosphate precipitation. The rate of this conversion is dependent on pH, temperature, and the presence of calcium hydroxyapatite seed crystals. (37)

The quantity of lime required to achieve a given pH level for calcium phosphate precipitation is a function of the wastewater alkalinity. Table 21 summarizes data taken at a number of locations to determine the lime requirements of the PhoStrip process. This data is plotted in Figure 25 along with the curve recommended in the EPA Process Design Manual for Phosphorus Removal. The reason for the variance between the EPA curve and the data presented in Table 21 is the lime dissolution efficiency. The data points to the right of the line were taken from a system which was not designed to precipitate phosphorus with high lime utilization as a goal. The scatter in the bench scale data is largely due to experimental error. This data does indicate good agreement with the correlation recommended by the EPA and thus this correlation should be used to determine the lime dose for the stripper supernatant from the PhoStrip process.

TABLE 21. SUMMARY OF LIME DOSAGE REQUIREMENT STUDIES OF THE STRIPPER SUPERNATANT

Location	Phase or Test	Alkalinity* mg $CaCO_3$/l	pH Initial	Final	Lime Dosage mg CaO/l
Tonawanda, N.Y.	I	310	7.0	9.1	225
(Pilot Scale)	II	310	6.7	10.5	380
	III	325	7.0	9.6	255
	IV	340	7.1	10.2	330
	V	370	7.0	10.1	424
	VI	380	6.2	9.2	312
Adrian, Mich.	1	210	6.9	9.0	93
(Bench Scale)	2	210	6.9	9.5	166
	3	215	6.7	9.1	141
	4	215	6.7	9.5	150
	5	235	6.9	9.1	133
Brockton, Mass.	1	165	7.2	9.3	64
(Bench Scale)	2	165	7.2	9.7	95
Texas City, Texas	1	350	7.2	9.0	123
(Bench Scale)	2	350	7.2	9.5	180

* Sum of carbonate plus phosphate alkalinity

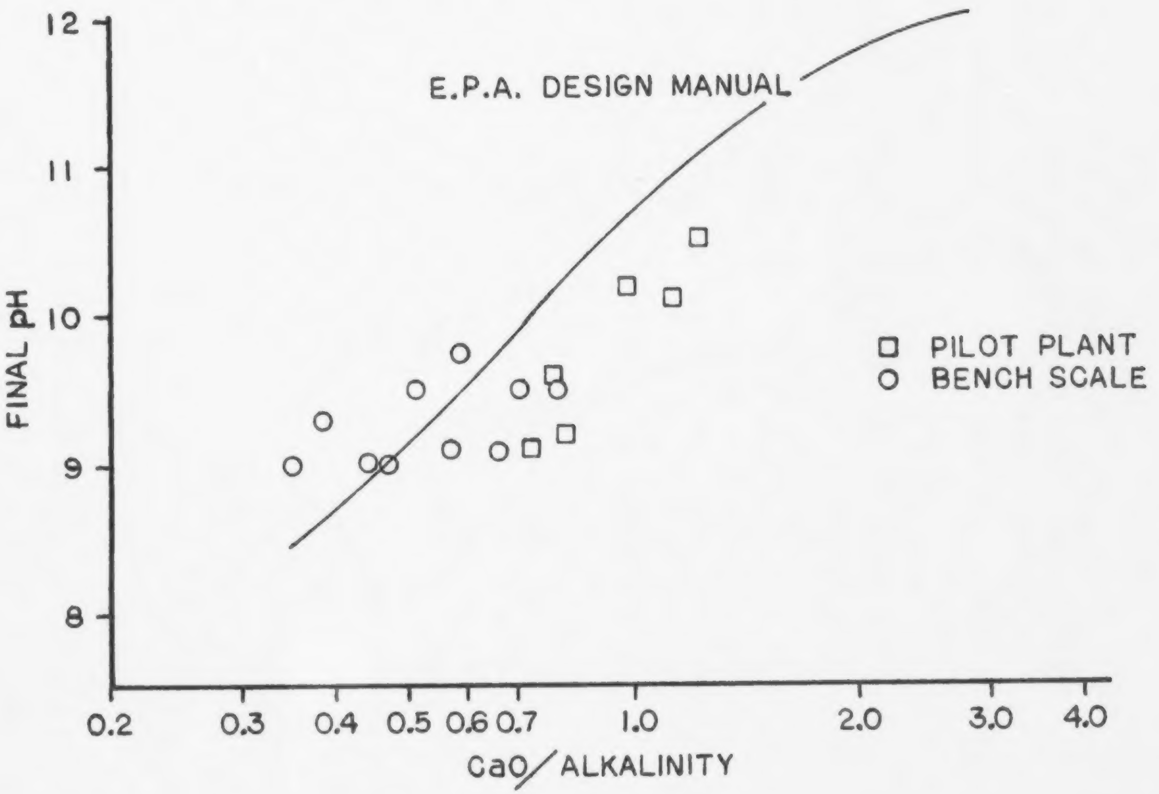

Figure 25. Lime requirements to achieve a given pH level.

As with the lime dosage, the amount of sludge produced via precipitation depends on the ability to efficiently dissolve lime. Sludge production data obtained for the bench scale studies performed at Adrian, Michigan and Texas City, Texas are presented in Table 22. This data demonstrates the relationship between sludge production, pH, and phosphorus level. Generally, an increase in either pH or phosphorus level will increase sludge production per volume of supernatant. An increase in pH increases sludge production mainly due to the formation of greater quantities of calcium carbonate and magnesium hydroxide. Consequently, the phosphorus content of sludges produced at increasing pH levels is lowered. Increasing the phosphorus level in the supernatant yields higher sludge production rates per unit volume, at a constant pH, due to increased calcium phosphate precipitation products. Thus, the phosphorus content of the waste sludge increases. For a given mass of phosphorus to be precipitated, increased supernatant phosphorus levels decrease chemical sludge production.

Figure 26 presents a comparison between the sludge production quantities obtained in the tests shown in Table 22 and those obtained through the use of the method proposed in the EPA Process Design Manual on Phosphorus Removal. (34) This comparison indicates that the calculational method is a good approximation of the data. At pH levels greater than 10.0, the calculated and observed results agree quite well. At a pH of ~ 9.5, the calculated values were approximately 8% lower than the measured values while at a pH of ~ 9.0, the measured values were 15% higher than those obtained from the calculations.

As discussed earlier, the precipitation portion of the PhoStrip process should be operated at pH ≈9.0 to minimize lime requirements and chemical sludge production. In order to obtain the proper sludge production quantities, it is recommended that the method presented in the EPA Design Manual for Phosphorus Removal (34) be employed with a 15% correction factor to account for the inaccuracies that appear at a pH level of 9.0.

Table 23 summarizes the results of filter leaf studies run to determine the filterability of the chemical sludge from the PhoStrip process. The sludge for this study was obtained from a reactor clarifier system which was used to precipitate the supernatant from a PhoStrip system operating on Tonawanda, N.Y. wastewater. The sludge concentration obtained from the reactor clarifier and used in these tests ranged in TSS levels between 7 and 12%. The solids concentration obtainable from full scale systems will depend on operational characteristics of the unit designed, but should range between 3 and 12%.

The results in Table 23 indicate that a vacuum filter can be used to produce a 35% solids cake at filter loadings between approximately 8 and 16 lbs/hr/ft^2. These loadings were achieved without the use of chemical

TABLE 22. CHEMICAL SLUDGE PRODUCTION FROM STRIPPER SUPERNATANT

Location	Phase or Test	pH Final	Phosphorus Removed mg/l	Sludge Produced mg/l Supernatant	$Ca_5(PO_4)_3OH$* Content %
Tonawanda, N.Y.	I	9.1	18	380	27
(Pilot Scale)	II	10.5	35	600	31
	III	9.6	36	530	36
	IV	10.2	41	710	31
	V	10.1	57	650	47
	VI	9.2	75	560	72
Adrian, Mich.	1	9.0	36	240	80
	2	9.5	36	320	61
	3	9.1	38	310	66
	4	9.5	38	340	60
	5	9.1	48	350	83
Texas City, Texas	1	9.0	26	210	67
	2	9.5	28	320	48

* calculated from phosphorus removed

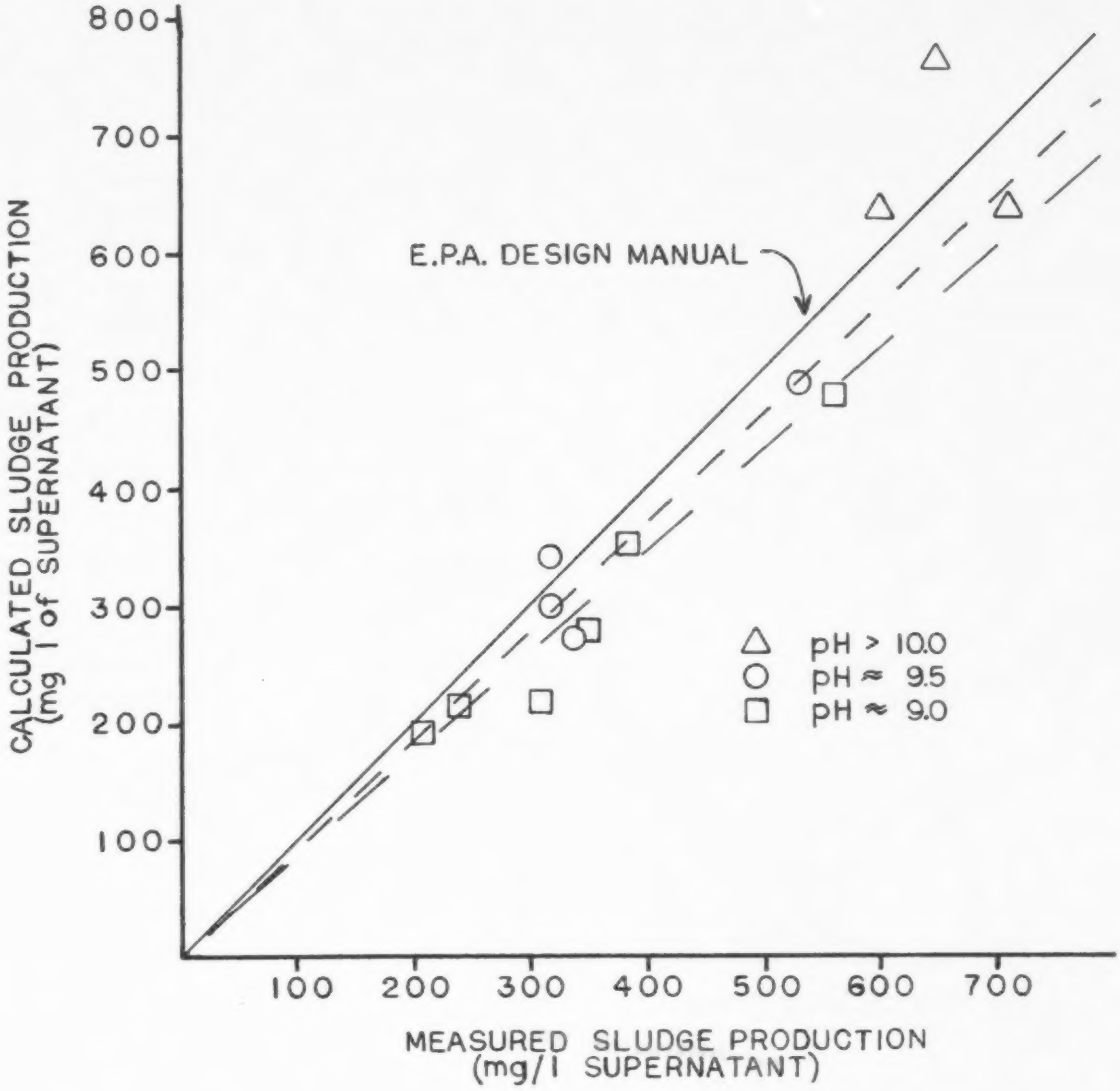

Figure 26. Comparison of calculated(34) and measured sludge production.

TABLE 23. FILTER LEAF TEST RESULT OF CALCIUM PHOSPHATE SLUDGE FORMED FROM STRIPPER SUPERNATANT

Test #	TSS Sludge %	Form Time sec.	Dry Time sec.	Cake Solids %	TSS Filtrate mg/l	Loading Rate lb/hr/ft^2*
1	11.5	30	50	33.9	98	15.9
2	11.5	45	75	34.5	51	11.9
3	11.5	45	105	35.4	39	9.4
4	11.5	45	135	36.3	159	7.8
5	11.5	60	100	35.5	86	10.3
6	11.5	30	100	37.2	117	9.2
7	11.5	45	105	35.9	99	12.3
8	7.3	30	50	32.5	52	9.5
9	7.3	45	75	32.8	43	7.9
10	7.3	45	75	32.7	30	8.9

Filter Cloth: Eimco NY529F

pH of quick mix tank = 9.1

* x 4.88 = kg/hr/m^2

conditioners. The filtrate produced in these tests were very low in suspended solids indicating an average solids capture in excess of 99%. As with most sludges, the loading capacity of the vacuum filter increases with increased suspended solids levels in the sludge to be dewatered.

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APPENDIX

(ECONOMIC EVALUATIONS)

These two case studies are verbatim excerpts from design reports. They have been included with the permission of the consulting firms involved and are for illustrative purposes only.

ADRIAN, MICHIGAN

The following has been extracted from "Professional Judgment Finding on 'PhoStrip' Phosphorus Removal System, Lenawee County Drain Commissioner for the City of Adrian, Michigan, Wastewater System Improvements, C262449" prepared by McNamee, Porter and Seeley, Consulting Engineers, Ann Arbor, Michigan, June 1977. This report was submitted to and approved by the Bureau of Environmental Protection, Michigan Department of Natural Resources.

Background

The City of Adrian is currently served by a 5 mgd activated sludge plant which discharges to the South Branch of the River Raisin. The secondary plant has been in operation since 1938 with additions made in 1950 and 1967. At present, flows during peak months average near the design capacity. An expansion of the plant is proposed to increase capacity for inclusion of domestic and industrial flows from Adrian and Madison Township surrounding the City. Included in the expansion is the addition of the capability of removing phosphorus in compliance with proposed final effluent discharge requirements.

In conjunction with the Step 2 design for expansion of the Adrian Wastewater Treatment Plant an investigation was made of available processes for phosphorus removal. Because of potential operation and maintenance cost savings and successful test operation at other locations, a pilot-scale operations study was begun in July 1976 to evaluate the applicability of the PhoStrip process for phosphorus removal. The results of the pilot study were presented to the Michigan DNR in December 1976 and clearly showed that the process effectively removed phosphorus from the Adrian wastewater in excess of the requirements of the proposed NPDES Permit.

General Process Description

The PhoStrip process utilizes activated sludge microorganisms to concentrate phosphorus from the wastewater flow stream into a relatively small substream from which removal by chemical precipitation is relatively inexpensive. The process takes advantage of the biological phenomena of "luxury uptake" and "anaerobiosis" for release of phosphorus.

A holding tank (stripping tank) is used to detain a portion of the return activated sludge (RAS) under anaerobic conditions for phosphorus release and to separate a clear phosphorus-rich liquor (supernatant) from the RAS flow stream. The high phosphorus concentration supernatant is transferred to a rapid mix tank where it is mixed with a lime slurry. The resulting mixture of supernatant, lime slurry and precipitated solids is then transferred to the primary clarifiers for separation by gravity sedimentation of the phosphorus precipitate.

The microorganisms take up phosphorus in the aeration tank and then are induced to release phosphorus while detained under anaerobic conditions in the stripping tank. The compacted activated sludge (anaerobic RAS) in the bottom of the stripping tank contains phosphorus deficient microorganisms and phosphorus rich liquor. The anaerobic RAS is returned to the aeration tank where the phosphorus uptake cycle is repeated. Anaerobic RAS is also recycled through the stripping tank by blending with the influent return activated sludge (aerobic RAS) from the 1st stage sedimentation tank in order to convey phosphorus-rich liquor to the surface of the stripping tank. The major difference between the PhoStrip process and other attempts to use biological processes for phosphorus removal is the use of a separate phosphorus stripping tank where phosphorus release is induced and controlled. Also, only a fraction of the RAS is diverted to the stripper (approximately 10-15% of the total design flow).

Justification for Use

The proposed Adrian WWTP design is for a two-stage activated sludge system. The first stage will be used for carbonaceous removal and the second stage will convert ammonia to nitrate. The PhoStrip process will be operated in conjunction with the first stage aeration system. The main advantage of the PhoStrip system is dosing only 10-15% of the sewage flow with chemicals as compared to the conventional method of chemical treatment which uses addition of chemical to the entire flow of sewage. Most of the operating expenses of phosphorus removal is the chemical cost and the PhoStrip system significantly reduces this cost. This will be shown in the cost-effective analysis in the following section.

Also, the PhoStrip process produces less solid waste which is another major cost savings, since solid wastes directly affect sludge handling and disposal operations.

Another advantage to the PhoStrip system is that it is less affected by shock load, as would be expected, because of the offstream reservoir of activated sludge provided by the stripping tank. This is especially important in Adrian's case because of the history of industrial pollutant shock loads that have been experienced in the raw sewage flow.

The pilot plant test data on the typical Adrian domestic wastewater conclusively demonstrated the satisfactory performance of the PhoStrip process for achieving greater than 90% removal of phosphorus.

The use of lime addition for phosphorus removal also enhances the process. The cost per ton of lime versus the traditional chemicals is less. Also, the nitrification stage of the plant will require pH adjustment to achieve complete nitrification due to low influent alkalinity. With a lime handling facility already provided for in the PhoStrip process, the system can also be used for the pH control needed throughout the plant. The lime-phosphorus sludge which will be transferred to the primary sedimentation tank will enhance the settleability of the raw sludge, thus increasing the removal efficiency of the tanks.

Cost Effective Analysis

An economic evaluation was made comparing PhoStrip against the traditional chemical addition methods using ferric chloride and alum. The cost-effective analysis included initial installment cost and total annual costs which includes capital cost amortized over 20 years at 6-1/8% interest; chemical cost; operating labor; maintenance and repair costs; and sludge disposal costs.

The PhoStrip costs were based on treatment of 15% of the maximum plant flow of 7 mgd, or 1.0 mgd. The cost of the two traditional chemical processes were based on total plant flow of 7 mgd. The bench scale lime dosage test conducted during the pilot plant operation established a lime dosage of 250 mg/l to produce the desired treatment. The dosage required for treatment with the traditional chemicals were: 90 mg/l of ferric chloride and 135 mg/l of alum.

The estimated total annual cost of increasing the capacity of the existing plant and adding the capability for phosphorus removal are: $85,565 per year using the PhoStrip process; $129,163 per year using the Ferric Chloride process; and $136,183 per year using the alum process.

Table EE-1
Cost Comparison Between PhoStrip and Chemical Addition for Phosphorus Removal at Adrian, Michigan

Design Flow = 7 mgd
Influent P = 10 mgd
Effluent P = 1 mg/l

Item	PhoStrip	Ferric Chloride	Alum
A. Installed Investment[1]	$520,000	$ 60,000	$ 65,000
B. Annual Costs			
1. Amortized Investment[2] (A x 0.08897)	46,265	5,338	5,783
2. Chemical Costs			
a. Lime[3]	20,000	---	---
b. Ferric Chloride[4]	---	105,500	---
c. Alum[5]	---	---	115,000
3. Operating Labor, Maintenance & Repair	8,000	3,000	3,500
4. Sludge Disposal Cost O&M	@565 T/yr	@765 T/yr	@770 T/yr
a. Anaerobic Digestion at $5/ton	2,800	3,825	3,850
b. Transport of Liquid Sludge by Tank Truck at $15/Ton	8,500	11,500	11,550
TOTAL ANNUAL COST	$ 85,565	$129,163	$136,183
$/mg	$33.50	$50.55	$53.30

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[1] See Table EE-2 for cost breakdown.

[2] Assumes a 20-year equipment life and 6-1/8% capital cost.

[3] Based on a 250 mg/l lime dosage, 15% Q_t supernate flow, and a lime cost of $50/ton delivered.

[4] Based on a 90 mg/l $FeCl_3$ dosage, 7 mgd flow, and a $FeCl_3$ cost of $110/ton delivered.

[5] Based on a 135 mg/l alum dosage, 7 mgd flow, and an alum cost of $80/ton delivered.

Table EE-2
PhoStrip Cost Comparison

Equipment Cost (PhoStrip)	Cost
1. Stripper Tank	
a. Concrete 65-ft. dia. at 20-ft. SWD	$150,000
b. Mechanism & Warranty (supplied by Union Carbide)	245,000
2. Lime-Mix Tank	
a. Concrete 9-sq. ft. at 12-ft. SWD	57,500
b. Mixer	2,500
3. Pumps	
a. Stripper supernatant Pumps	2,000
b. Anaerobic RAS Pumps	2,000
c. Stripper Waste Pumps	1,000
4. Lime Feed Equipment* (50 ton Bin, Feeder, Slaker)	60,000
Total Installed Cost	$520,000

* Cost taken from EPA Manual "Phosphorus Removal", p. 10-32

$$\text{ENR} = \frac{2520_{(1977)}}{1643_{(1971)}} = 1.53$$

Equipment Cost (Ferric Chloride)	
1. Bulk Storage 2/8000 gal. tanks	$ 25,000
2. Pumps	
Transfer	15,000
Feed	
3. Dilution and Feed Tanks, Agitation, Piping	20,000
Total Installed Cost	$ 60,000

Equipment Cost (Alum)	
1. Bulk Storage 4/8000 gal. tanks	$ 50,000
2. Pumps	
Transfer	15,000
Feed	
Metering	
Total Installed Cost	$ 65,000

The savings in total annual costs through the use of the PhoStrip process amount to approximately $44,000 per year, chemical costs showing the largest cost savings. The cost per million gallons of the traditional chemical addition is approximately twice as much as the PhoStrip process. The PhoStrip process requires a higher initial investment (approximately $460,000 or $40,000 per year amortized cost difference).

SOUTHTOWNS, NEW YORK

The following was extracted from an addendum to the Facilities Report for the Southtowns Sewage Treatment Agency, Erie County, New York, prepared by McPhee, Smith, Rosenstein Engineers, Montvale, New Jersey. This report was submitted to and approved by the New York State Department of Environmental Conservation.

Phosphate Removal

1) General

As previously described, the present plans call for the removal of phosphorus from the wastewater by post precipitation of the biologically treated effluent, followed by filtration. The cost of alum to remove the phosphorus will amount to a sizable portion of the operating budget of the facilities. The PhoStrip process can accomplish the phosphorus removals at a lower capital and operating cost.

Messrs. Levin, Topol, Tarnay and Samworth reported pilot plant work on this new process for the removal of phosphates. The PhoStrip process is a unique system that accomplished phosphorus removal by dosing only a small fraction of the sewage flow with chemicals instead of dosing the total flow.

The PhoStrip process is based on three characteristics experienced in sewage treatment:

a) The activated sludge microorganisms in the mixed liquor of an activated sludge process can be induced to take up soluble phosphorus. This is known as "luxury uptake" since the quantity "taken up" is in excess of the amount required for growth.

b) If the mixed liquor solids are denied oxygen and once all the dissolved oxygen available is consumed, the microorganisms will release the phosphorus previously taken up, and it will go back into solution.

c) If the supernatant of the mixed liquor solids is treated with lime similar to the cold lime softening of water, the soluble P can be precipitated and removed from the process.

The PhoStrip process accomplishes the removal of phosphorus by utilizing the above characteristics in the following manner (see Figure EE-1):

a) A fraction of the sludge from the bioclarifier underflow is conveyed to an anaerobic phosphate stripper. The remaining fraction is returned to the oxygen reactors. The dilution time in this stripper is set so that the contents will consume all the dissolved oxygen present and pass into an anaerobic state. As the microorganisms give up the soluble P they had taken up, they are washed in the stripper by an upflow of low P effluent water from the phosphorus precipitator (reactor clarifier). An alternate source of the elutriation stream will be the effluent from the biological clarifier.

b) The phosphorus-enriched stripper supernatant flows to a solids-contact clarifier where lime is added to precipitate the phosphorus. As stated previously, the effluent from the clarifier is returned as an elutriation stream for the stripper. The sludge from the clarifier will be pumped to a holding tank prior to dewatering on a vacuum filter.

As can be noted from the above description, only a small percentage of the total plant flow has to be treated to remove the necessary amount of phosphorus. The volume of the supernatant stream will be approximately 10% of the total plant flow, thereby reducing the size of the treatment units and the quantity and cost of the chemicals removed.

c) The phosphorus stripped sludge is withdrawn from the bottom of the stripper and returned to the oxygen reactors. The sludge has a relatively large capacity to pick up soluble phosphorus from the wastewater.

2) Pilot Plant Data

Laboratory pilot plant work was done by Levin et al and was reported in the October 1974 Journal, Water Pollution Control Federation. This report concluded:

a) that the new process would remove 90% of the total phosphorus present in the wastewater;

b) by the addition of the filtration step, the removal could be increased to 95%;

c) the method was compatible with the conventional activated sludge process;

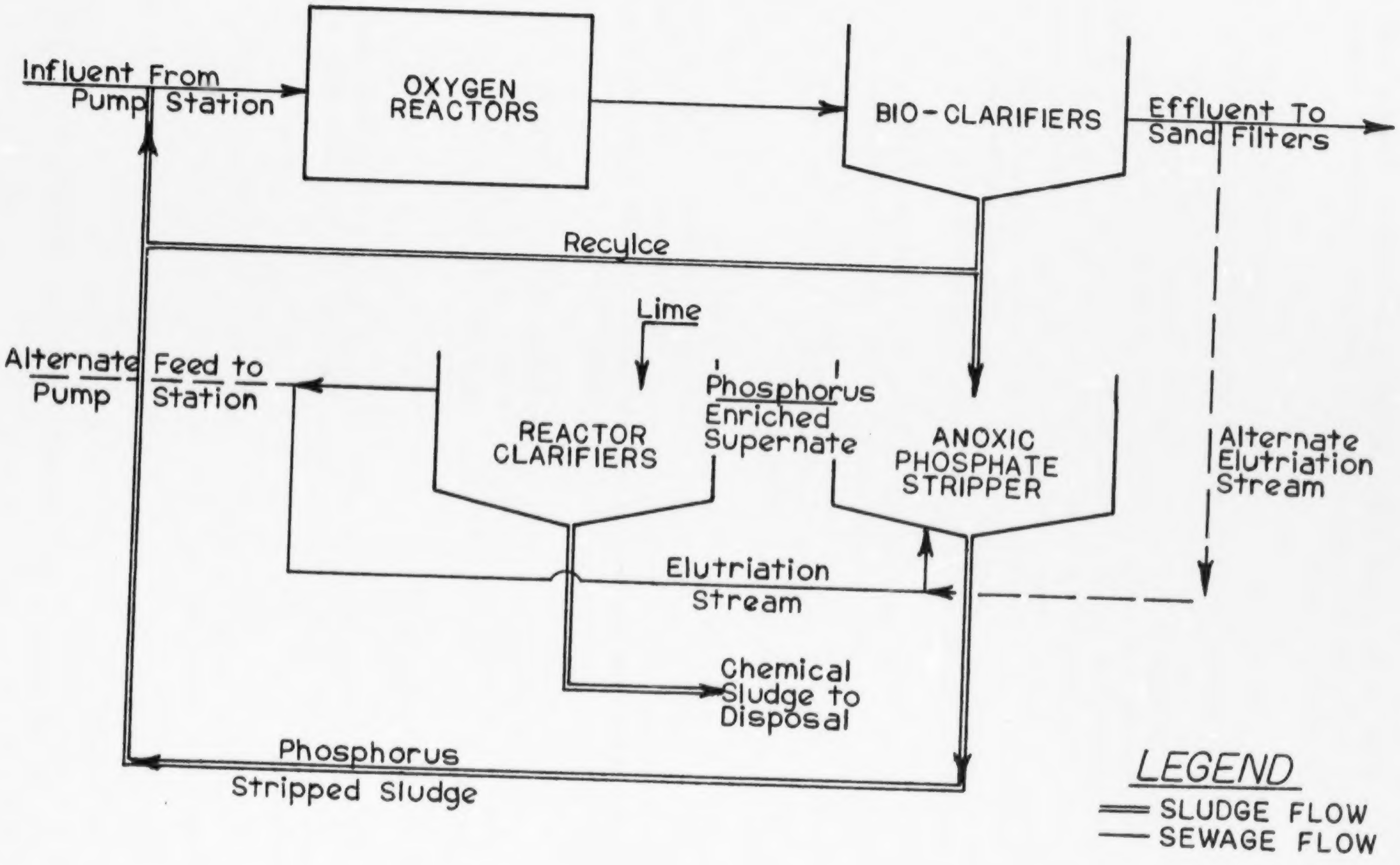
Influent From Pump Station
OXYGEN REACTORS
BIO-CLARIFIERS
Effluent To Sand Filters
Recylce
Lime
Alternate Feed to Pump Station
REACTOR CLARIFIERS
Phosphorus Enriched Supernate
ANOXIC PHOSPHATE STRIPPER
Alternate Elutriation Stream
Elutriation Stream
Chemical Sludge to Disposal
Phosphorus Stripped Sludge
LEGEND
SLUDGE FLOW
SEWAGE FLOW

Figure EE-1. PhoStrip flow pattern.

d) the sludge produced was less than produced if the entire waste stream were treated with lime;

e) activated sludge process was improved; and

f) the method provided a reserve reservoir of mixed liquor suspended solids.

The City of Reno, Nevada instituted a full-scale plant size study. The work was begun in May 1974 and was reported in a paper given at the 48th Annual Conference of the Water Pollution Control Federation on October 7, 1975. The paper concluded:

a) that in a 6 mgd plant scale test, 90% of the phosphorus could be removed;

b) the process could be operated with relative ease;

c) a saving of $600,000 to $800,000 a year could be realized in the O&M of a 40 mgd plant.

After Union Carbide purchased the PhoStrip process, they continued to test the process and improve on it. Their work has improved the process as far as reducing its capital costs and operating costs are concerned. They ran a pilot plant at their Tonawanda, N.Y. facilities, and Table EE-3 summarizes the data from the Tonawanda pilot plant for five different phases. The phases tried are as follows:

I) No elutriation of the sludge in the anaerobic stripper.

II, III, IV) Effluent from a primary clarifier was recycled as an elutriant.

V) The effluent from a solid contact clarifier was recycled as an elutriant.

The table shows that the average total P in the effluent for all test phases was 1.0 mg/l or less.

Additional pilot plant work was done at Texas City, Texas. A 10 gph pilot plant, consisting of primary clarifier, aeration basins, secondary clarifier and anaerobic stripper was used. Table EE-4 shows the results of this plant for a 14-day steady state phase. The average total P in the effluent was measured as 0.7 mg/l and the maximum of 1.5 mg/l. Since the plant did not have a continuous lime precipitation phase, the results are based on jar tests of the effluent.

Table EE-3
PhoStrip System Pilot Plant Testing

	I	II	III	IV	V
Feed Rate, GPM	12.1	10.0	13.0	15.0	15.0
% of total flow as:					
Sludge recycle direct to aeration*				20.7	14.0
Sludge feed to stripper	24.8	33.5	23.7	19.4	20.0
Supernatant from stripper	13.2	19.3	14.9	9.1	11.9
Temperature, F	56	57	61.2	64.4	72
pH, mixed liquor	7.0	6.7	6.7	6.7	6.4
pH, stripper supernatant	6.7	7.0	7.1	7.0	6.3
MLSS, mg/1	2930	4820	3620	5190	4750
MLVSS, mg/1	2320	3600	2710	3750	3490
F/M (organic loading),					
lb. BOD_5app/day/lb.MLVSS	0.51	0.27	0.60	0.38	0.44
Retention time-aeration, hr.	1.6	2.5	1.9	1.7	1.7
Retention time-stripper, hr.	9.9	5.4	5.9	6.1	6.4
Phosphorus concentration, mg/1					
Influent	5.3	6.8	6.6	7.5	10.5
Effluent (total P)	0.9	0.4	0.5	0.5	0.9
Effluent (soluble P)	0.7	0.2	0.3	0.3	0.4
Stripper supernatant	27.0	38.0	43.0	53.5	70.5
Stripper underflow (filtered)	59.0	22.0	34.0	36.5	46.0
Clarifier overflow rate, gpd/ft^2	580	480	630	725	725
Lb.P/lb.MLVSS	0.027	0.024	0.028	0.033	0.040
Effluent quality					
ESS,mg/1	10	7	10	7	10
EVSS, mg/1	9	7	6	5	8
BOD, influent, mg/1	78	96	139	100	110
BOD, total effluent, mg/1	10	8	9	10	13
Duration of test, days	10	12	29	29	37

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* All sludge passed through the stripper during test Phase I, II and III before recycle to aeration.

Table EE-4
PhoStrip Pilot Plant - Texas City, Texas

Phosphorus Balance Data - Steady State Phase

	INFLUENT		EFFLUENT		STRIPPER SUPERNATE		FILTERED AERATION EFFLUENT		STRIPPER INF	WASTE SLUDGE	STRIPPER UNDERFLOW	
Date	Total P mg/1	Ortho mg/1	Total P mg/1	Ortho P mg/1	Total P mg/1	Ortho P mg/1	Total P mg/1	Ortho P mg/1	Total P mg/1	Total P mg/1	Total P mg/1	Filtered Ortho P mg/1
3-17-76	7.6	7.0	0.4	0.4	144	52	0.2	0.1	520	364	552	25
3-18-76	8.0	6.2	0.8	0.3	112	50	0.2	0.1	332	320	288	22
3-19-76	9.0	7.0	1.4	0.8	36	42.5	0.1	0.1	264	-	288	18.5
3-20-76	8.4	5.2	0.2	0.1	29	34.0	5.8	6.2	288	124	332	42.5
3-21-76	6.4	3.8	0.5	0.0	27.5	33.0	3.8	4.0	364	320	312	48.0
3-22-76	4.6	4.2	0.4	0.0	29.0	34.0	0.4	0.0	280	364	332	40.0
3-23-76	-	6.6	0.5	0.5	53.0	38.5	0.2	0.4	304	140	264	27.0
3-24-76	11.0	6.0	1.5	0.5	45	37.5	1.8	1.2	352	312	296	22.5
3-25-76	8.3	5.1	0.6	0.2	30	23	0.6	0.4	280	220	256	18.0
3-26-76	7.6	6.0	0.9	0.7	61.5	35	0.9	0.8	264	192	232	13.5
3-27-76	7.6	5.7	0.2	0.2	34	31	-	-	312	432	296	61.0
3-28-76	-	6.6	0.6	0.2	18	28.5	-	3.7	200	-	388	70
3-29-76	9.1	6.2	0.6	0.4	24	28.5	0.9	0.3	388	388	272	40
3-30-76	6.8	5.4	0.6	0.2	22	25	0.1	0.1	408	-	364	44.5
AVERAGE	7.9	5.8	0.7	0.3	47.5	35.9	1.3	1.3	325	289	319	35.2
MAXIMUM	11.0	7.0	1.5	0.8	144	52	3.8	6.2	520	432	552	70
MINIMUM	4.6	3.8	0.2	0.0	18	23	0.1	0.0	200	124	232	13.5

A nine-week pilot study of waste from the Hamburg Master District has just been concluded. The average effluent concentration of soluble P in the effluent was 0.1 mg/l, well below the design value of 0.5 mg/l. The average total P in the effluent was 1.4 mg/l. These effluent concentrations are following clarification. There weren't any filters in the pilot plant. The effluent solids averaged 58 mg/l, but this was because of the small scale of the pilot plant clarifier. In a full scale operation, the effluent concentration of total P and suspended solids should be lower. In addition, the Southtowns treatment plant will have filtration following clarification, which will result in lower effluent concentrations of P and SS.

3) Cost Effective Analysis

An economic analysis of three (3) design alternates follows. The first alternate is the present design which consists of using aluminum salts for phosphate removal, incineration of a combination of chemical and waste activated sludge and trucking the ash to a landfill. Alternate II is the PhoStrip process which substitutes quicklime for alum. For this alternate the PhoStrip sludge would be dewatered separately, the waste activated sludge would be incinerated and a combination of PhoStrip sludge and incinerator ash would be hauled to a landfill. The third alternate consists of the incineration of the PhoStrip sludges along with waste activated sludges. The ash would then be hauled to a landfill. In addition, the on-site disposal of the residue for the three alternates was studied. These are referred to as Alternates IA, IIA, and IIIA.

Tables EE-5 and EE-6 present the annual and operating costs for the initial and the 20th year of operation. The cost for dumping sludge in a segregated landfill at the Lancaster site has been assumed to be $10.60/ton. The calculations in Tables EE-5 and EE-6 are based on the assumption that a 15 cubic yard truck will be used to haul sludge to Lancaster. Chemical costs of $30/ton for quicklime, $95/ton for Al_2O_3 and $130/ton for ferric chloride have been assumed. Fuel oil is assumed to cost $0.30/gallon.

Table EE-7 summarizes the capital costs associated with the six alternates. The capital cost of the PhoStrip equipment is approximately $2,100,000 versus $2,650,000 for the present system. The capital cost for the PhoStrip system with the sludge being incinerated will be approximately $2,200,000. The engineering costs to revise the plans and specifications to incorporate the PhoStrip process will cost $210,000. It has been assumed that this design will delay the project a maximum of four (4) months until all Federal and State approvals are obtained.

The outfall sewer will be bid on time. The escalation in the other contracts will be 6%. It is assumed that the contractor costs will escalate approximately ($28,000,000)(.06) 4/12 = $560,000. If the residue

Table EE-5
Annual Costs - Initial Year

	Alt. I Alum Sludge		Alt. II PhoStrip		Alt. III PhoStrip/Incin.	
	To Lancaster	On Site	To Lancaster	On Site	To Lancaster	On Site
Ash or Ash & Sludge						
#/d	6,108	6,108	12,353	12,353	6,444	6,444
#/CF	150	150	105	105	150	150
CY/day	1.5	1.5	4.4	4.4	1.6	1.6
Trips/week	1	1	2	2	1	1
Trucking Cost ($/yr)						
Depreciation	4,000	4,000	4,000	4,000	4,000	4,000
Maintenance	1,000	1,000	1,000	1,000	1,000	1,000
Gas/Oil	1,300	100	2,600	200	1,300	100
Labor $16,000/yr.	1,600	200	3,200	400	1,600	200
Sub Total	7,900	5,300	10,800	5,600	7,900	5,300
Annual Cost						
Al_2O_3 or CaO	39,182	39,182	8,020	8,020	8,020	8,020
$FeCl_3$	31,505	31,505	26,139	26,139	33,240	33,240
CaO	3,635	3,635	3,016	3,016	3,835	3,835
Add'tl Fuel Oil/Incinerator	-	-	-	-	2,208	2,208
Landfill Charge	11,815	-	23,896	-	12,465	-
Sub Total	86,137	74,322	61,071	37,175	59,768	47,303
Sub Total	7,900	5,300	10,800	5,600	7,900	5,300
TOTAL	$94,037	$79,622	$71,871	$42,775	$67,668	$52,603

Table EE-6
Annual Costs - Year-20

	Alt. I Alum Sludge		Alt II PhoStrip		Alt. III PhoStrip/Incin.	
	To Lancaster	On Site	To Lancaster	On Site	To Lancaster	On Site
Ash or Ash & Sludge						
#/d	13,783	13,783	27,873	27,873	14,542	14,542
#/CF	150	150	105	105	150	150
CY/day	3.4	3.4	9.8	9.8	3.6	3.6
Trips/week	2	2	5	5	2	2
Trucking Cost ($/yr)						
Depreciation	4,000	4,000	4,000	4,000	4,000	4,000
Maintenance	1,000	1,000	1,000	1,000	1,000	1,000
Gas/Oil	2,600	200	6,500	500	2,600	200
Labor $16,000/yr	3,200	400	8,000	1,000	3,200	400
Sub Total	10,800	5,600	19,500	6,500	10,800	5,600
Annual Cost						
Al_2O_3 or CaO	88,507	88,507	18,094	18,094	18,094	18,094
$FeCl_3$	71,089	71,089	58,980	58,980	75,004	75,004
CaO	8,202	8,202	6,805	6,805	8,654	8,654
Add'tl Fuel Oil/Incinerator	-	-	-	-	4,990	4,990
Landfill Charge	26,663	-	53,920	-	28,131	-
Sub Total	194,461	157,798	137,799	83,879	134,873	106,742
Sub Total	10,800	5,600	19,500	6,500	10,800	5,600
TOTAL	$205,261	$173,398	$157,299	$90,379	$145,673	$112,342

Table EE-7
Capital Costs

	Alt. I		Alt. II		Alt. III	
	To Lancaster	On Site	To Lancaster	On Site	To Lancaster	On Site
Construction Cost	2,650,000	2,650,000	2,100,000	2,100,000	2,200,000	2,200,000
Phosphorus Removal Sect. Development of Landfill at Site	-	100,000	-	100,000	-	100,000
Engineering Cost of Revising Plans & Specs	-	-	210,000	210,000	210,000	210,000
Additional Demolition Due to Landfill at Site	200,000	200,000	200,000	200,000	200,000	200,000
Inflation on Treatment Plant, 4 mo. at 6%	-	-	560,000	560,000	560,000	560,000
	$2,850,000	$2,950,000	$3,070,000	$3,170,000	$3,170,000	$3,270,000
- Salvage Value	265,000	265,000	210,000	210,000	220,000	220,000
TOTAL	$2,585,000	$2,685,000	$2,860,000	$2,960,000	$2,950,000	$3,050,000

is to be disposed of on the existing site, it has been estimated that it will cost $100,000 to develop a segregated landfill at the site. Erie County was planning to demolish all the silos and buildings on the site, even those that would not have been eligible for Federal and State aid. If the residue is disposed of on the site, the additional demolition work that wasn't necessary in order to build treatment facilities would not be necessary in order to develop an on-site landfill. The estimated cost for this additional demolition ($200,000) will not be eligible for Federal and State aid.

A cost-effective analysis based on a 20-year planning period at 6-1/8% interest for the six (6) alternates is shown in Table EE-8.

4) Conclusions and Recommendations

A summary of the present worth of the six (6) alternates is given below:

Alternate	Present Worth
I	$4,131,000
IA	$3,990,000
II	$4,041,000
IIA	$3,649,000
III	$4,051,000
IIIA	$3,902,000

The cost-effective analysis for the 20-year period shows that Alternate IIA, the PhoStrip process with the chemical sludge dewatered separately and disposed of on the site along with the incinerator ash is the most cost-effective method. It is our recommendation that this process be specified as the exclusive process for the Erie County/Southtowns Wastewater Treatment Plant. We recommend that the rapid sand filters as planned for the original design be used to assure that the phosphorus in the effluent is consistently less than 1.0 mg/l as P. The insoluble phosphates in the bioclarifier effluent, along with the phosphorus tied up in the suspended solids, will be removed by the sand filters.

Table EE-8
Present Worth Analysis

ALTERNATE I - Alum Sludge to Lancaster

Present Worth of Annual Costs = (94,037) (11.3549) = 1,067,780

$\frac{(205,261-94,037)}{20}$ (85.93) = 477,873

Capital Costs = 2,585,000

Total Present Worth $4,130,653

SAY $4,131,000

ALTERNATE IA - Alum Sludge on Site Disposal

Present Worth of Annual Costs = (79,622) (11.3549) = 904,099

$\frac{(173,398-79,622)}{20}$ (85.39) = 400,376

Capital Costs = 2,685,000

Total Present Worth $3,989,475

SAY $3,990,000

ALTERNATE II - PhoStrip to Lancaster

Present Worth of Annual Costs = (71,871) (11.3549) = 816,088

$\frac{(157,299-71,871)}{20}$ (85.39) = 364,734

Capital Costs = 2,860,000

Total Present Worth $4,040,822

SAY $4,041,000

(Continued)

Table EE-8
Present Worth Analysis
(Continued)

ALTERNATE IIA - PhoStrip on Site

Present Worth of Annual Costs	= (42,775) (11.3549) =	485,705
	$\frac{(90,379-42,775)}{20}$ (85.39) =	203,245
Capital Costs =		2,960,000
	Total Present Worth	$3,648,950
	SAY	$3,649,000

ALTERNATE III - PhoStrip - Incinerator - to Lancaster

Present Worth of Annual Costs	= (67,668) (11.3549) =	768,363
	$\frac{(145,673-67,668)}{20}$ (85.39) =	333,042
Capital Costs =		2,950,000
	Total Present Worth	$4,051,405
	SAY	$4,051,000

ALTERNATE IIIA - PhoStrip - Incinerator - On Site Disposal

Present Worth of Annual Costs	= (52,603) (11.3549) =	597,301
	$\frac{(112,342-52,603)}{20}$ (85.39) =	255,055
Capital Costs =		3,050,000
	Total Present Worth	$3,902,356
	SAY	$3,902,000

APPENDIX A

(RAW DATA SUMMARY)

English to SI Conversions

GPM x 6.3 x 10^{-5} = m^3/s

in. x .0254 = m

(T °F - 32) x 0.555 = T °C

GAL x 3.785 x 10^{-3} = m^3

MGD x 3785 = m^3/D

(FT^3/GAL) x 7.481 = m^3/m^3

TABLE A-1. PHASE I OPERATING CONDITIONS FOR PHOSTRIP PILOT PLANT - RENO/SPARKS, NEVADA

DATE	PLANT INF GPM	SLUDGE RECIRCULATION		STRIPPER INFLUENT*		STRIPPER UNDERFLOW		AER RT_Q HRS	AVG D.O. mg/1	STRIPPER BLANKET HEIGHT INCHES	STRIPPER SRT HRS	M.L. TEMP °F	MLSS mg/1	MLVSS mg/1	STRIPPER* INFLUENT		STRIPPER UNDERFLOW		WASTE SLUDGE GAL/DAY
		GPM	%Q	GPM	%Q	GPM	%Q								RSS mg/1	RVSS mg/1	RSS mg/1	RVSS mg/1	
11-19-75	5.0	0.5	10	1.25	25	0.5	10	4.9	0.6	46.5	10.2	59	1796	1248	6648	5664	17344	14696	150
11-20-75	5.0	0.5	10	1.25	25	0.5	10	4.9	0.6	43.5	9.6	60	1388	1228	6468	5468	16020	13556	150
11-21-75	5.0	0.5	10	1.25	25	0.5	10	4.9	0.6	42	9.3	60	1560	1332	6360	5404	14556	12360	150
11-22-75	5.0	0.5	10	1.25	25	0.5	10	4.9	3.2	36	7.9	59	1340	1204	6360	5476	17164	14240	112.5
11-23-75	5.0	0.5	10	1.25	25	0.5	10	4.9	2.2	42 .	9.3	60	1688	1496	7284	6192	15640	13064	131.3
11-24-75	5.0	0.5	10	1.25	25	0.51	10	4.9	3.0	41.5	9.0	61	1208	1120	5590	4910	11810	10000	150
11-25-75	5.0	0.5	10	1.25	25	0.78	16	4.9	1.9	38	5.4	61	1000	880	5180	4472	10612	9056	112.5
11-26-75	5.0	0.5	10	1.25	25	0.80	16	4.9	1.3	33.5	4.6	61	1114	894	6510	5476	8064	6770	75
11-27-75	5.0	0.5	10	1.25	25	0.77	15	4.9	1.7	36	5.2	62	1407	1127	7911	6677	9485	8011	18.8
11-28-75	5.0	0.5	10	1.25	25	0.78	16	4.9	1.8	42	5.9	60	1654	1301	6800	5570	10850	9070	50
11-29-75	5.0	0.5	10	1.25	25	0.77	15	4.9	2.9	49	7.0	57	1414	1154	7510	6120	10720	9030	75
11-30-75	5.0	0.5	10	1.25	25	0.77	15	4.9	2.6	48	6.9	60	1596	1380	6790	5643	11032	9278	82.5
12-01-75	5.0	0.5	10	1.25	25	0.78	16	4.9	1.2	50	7.1	61	1194	1161	6070	5336	9838	8618	20
12-02-75	5.0	0.5	10	1.25	25	0.78	16	4.9	1.6	50	7.1	60	1430	1150	6960	5910	10790	9180	18.8
12-03-75	5.0	0.5	10	1.25	25	0.78	16	4.9	0.8	55	7.8	61	1150	-	6530	5880	11800	10370	67.5
12-04-75	5.0	0.5	10	1.25	25	0.78	16	4.9	1.7	51.5	7.3	62	1550	1340	7020	6070	11639	3335	47.5
12-05-75	5.0	0.5	10	1.25	25	0.65	13	4.9	2.2	56.5	9.6	63	1590	1480	7130	6270	11310	10820	60
12-06-75	5.0	0.5	10	1.25	25	0.48	10	4.9	1.8	57.5	13.2	61	1510	1470	5090	4580	13170	11450	35
12-07-75	5.0	0.58	12	1.24	25	0.49	10	4.9	-	64	14.4	59	700	680	4590	4010	11200	9660	50
12-08-75	5.0	0.59	12	1.25	25	0.52	10	4.9	2.6	56	11.9	60	1156	1108	4780	4240	14660	12700	55
AVERAGE	5.0	0.51	10	1.25	25	0.65	13	4.9	1.8	46.9	8.0	60	1372	1198	6379	5469	12385	10263	81.2

* SECONDARY CLARIFIER UNDERFLOW

TABLE A-2. PHASE I ORGANIC & SUSPENDED SOLIDS REMOVALS FOR PHOSTRIP PILOT PLANT RENO/SPARKS, NEVADA

DATE	COD, mg/1 INF	COD, mg/1 EFF	TSS, mg/1 INF	TSS, mg/1 EFF	VSS, mg/1 INF	VSS, mg/1 EFF
11-19-75	190	108	203	45	187	40
11-20-75	207	58	121	56	114	50
11-21-75	199	50	106	38	96	38
11-22-75	207	99	138	31	84	-
11-23-75	257	91	230	-	219	-
11-24-75	169	25	146	28	126	17
11-25-75	270	34	128	51	123	45
11-26-75	228	42	116	48	96	42
11-27-75	367	25	198	40	172	37
11-28-75	266	38	81	29	69	23
11-29-75	241	21	130	43	103	35
11-30-75	236	59	104	47	-	43
12-01-75	214	38	95	28	93	18
12-02-75	294	46	132	39	114	35
12-03-75	176	34	96	35	81	24
12-04-75	235	34	116	28	102	25
12-05-75	172	29	63	23	59	15
12-06-75	260	25	143	33	129	24
12-07-75	176	34	93	26	83	16
12-08-75	201	50	73	11	66	7
AVERAGE	228	47	126	36	111	30

TABLE A-3. PHASE I PHOSPHORUS BALANCE DATA FOR PHOSTRIP PILOT PLANT - RENO/SPARKS, NEVADA

DATE	INFLUENT		EFFLUENT		STRIPPER SUPERNATE		AERATION EFFLUENT		* STRIPPER INFLUENT	WASTE SLUDGE	STRIPPER UNDERFLOW		CLARIFIER UNDERFLOW FILTERED
	TOTAL P mg/1	ORTHO P mg/1	TOTAL P mg/1	ORTHO P mg/1	TOTAL P mg/1	ORTHO P mg/1	TOTAL P FILTERED mg/1	TOTAL P mg/1	TOTAL P mg/1	TOTAL P mg/1	TOTAL P mg/1	FILTERED TOTAL P mg/1	TOTAL P mg/1
11-19-75	5.5	3.7	1.9	.16	10.0	10.0	0.5	43	210	210	470	95	19
11-20-75	6.1	6.1	2.1	1.5	12.0	10.5	0.5	41	190	190	430	94	5.5
11-21-75	6.6	5.3	2.3	2.04	11.0	12.0	2.0	47	245	245	435	85	9.5
11-22-75	6.4	4.4	1.0	.64	9.0	10.0	0.5	44	235	235	540	89	4
11-23-75	7.0	5.0	2.5	2.64	11.0	12.0	2.0	46	230	230	480	79	8.5
11-24-75	6.2	4.5	3.3	3.48	9.2	12.0	2.4	48	240	240	470	85	10.5
11-25-75	8.6	7.0	4.7	4.8	10.1	13.0	3.4	37	205	205	385	76	9.5
11-26-75	8.2	6.6	4.6	3.85	10.8	12.0	2.8	50	270	270	335	47	13
11-27-75	9.5	6.6	4.3	4.5	10.0	13.0	2.4	59	300	300	340	49	17
11-28-75	7.8	6.5	4.4	4.4	10.3	26.0	2.6	57	290	290	400	51	16
11-29-75	8.3	5.9	2.8	1.95	10.5	8.0	3.0	60	270	270	370	51	12
11-30-75	8.2	7.2	4.4	4.1	11.2	13.0	3.6	59	280	280	390	62	13
12-01-75	9.2	8.6	4.9	4.7	14.6	19.0	4.1	57	265	265	365	63	18.5
12-02-75	10.1	9.4	5.7	5.5	16.2	19.0	5.7	61	260	260	380	73	-
12-03-75	6.7	5.1	4.2	4.5	21.0	30.0	3.5	52	250	250	410	79	17.5
12-04-75	8.8	7.2	3.7	3.1	25.0	28.0	3.0	58	260	260	385	86	-
12-05-75	7.9	5.7	3.2	2.7	25.0	25.0	2.2	55	265	265	400	92	-
12-06-75	9.3	6.8	2.8	2.65	17.5	18.0	1.5	57	200	200	395	102	10
12-07-75	9.6	8.4	4.0	3.2	13.5	15.0	3.0	40	190	190	415	95	15
12-08-75	9.7	8.2	3.8	3.2	13.5	15.0	2.8	41	185	185	460	94	6.5
AVERAGE	8.0	6.4	3.5	3.2	13.6	16.0	2.6	51	242	242	413	77	12
MAXIMUM	10.1	9.4	5.7	5.5	25.0	30.0	5.7	61	300	300	540	102	18.5
MINIMUM	5.5	3.7	1.0	.16	9.2	8.0	0.5	37	185	185	335	47	4

*SECONDARY CLARIFIER UNDERFLOW

TABLE A-4. PHASE II OPERATING CONDITIONS FOR PHOSTRIP PILOT PLANT - RENO/SPARKS, NEVADA

DATE	PLANT INF GPM	ELUTRIATION		STRIPPER* INFLUENT		STRIPPER UNDERFLOW		AER RT_Q HRS	AVG D.O. mg/1	STRIPPER BLANKET HEIGHT INCHES	STRIPPER SRT HRS	M.L. TEMP °F	MLSS mg/1	MLVSS mg/1	STRIPPER* INFLUENT		STRIPPER UNDERFLOW		WASTE SLUDGE GAL/DAY	CHEMICAL CLARIFIER INFLUENT pH
		GPM	%Q	GPM	%Q	GPM	%Q								RSS mg/1	RVSS mg/1	RSS mg/1	RVSS mg/1		
2-23-76	4.7	0.74	16	0.71	15	0.57	12	6.0	1.1	69	13.3	59	944	916	6860	-	8090	-	84	-
2-24-76	4.7	0.76	16	0.71	15	0.56	12	6.0	1.2	68	13.4	60	836	708	5090	4270	7570	6440	99	-
2-25-76	4.7	0.76	16	0.69	15	0.56	12	6.0	1.2	63	12.4	59	928	836	-	-	8290	7060	96	-
2-26-76	4.7	0.75	16	0.71	15	0.57	12	6.0	1.1	64	12.4	60	828	752	5970	5160	8540	7440	96	-
2-27-76	4.7	0.75	16	0.71	15	0.57	12	6.0	1.3	67.5	13.0	60	1004	876	5290	4530	7740	6760	106	-
2-28-76	-	-	-	-	-	-	-	-	2.0	67	-	61	866	758	5730	4710	7360	6150	118	-
2-29-76	-	-	-	0.72	15	-	-	-	1.8	66	-	60	812	662	5080	4340	6780	5800	122	8.2
3-01-76	4.7	0.75	16	0.69	15	0.6	13	6.0	2.5	56	10.3	59	802	684	5610	4900	7550	6290	96	8.8
3-02-87	4.7	0.72	15	0.70	15	0.55	12	6.0	2.3	59	11.8	58	964	856	5243	4630	8540	7320	102	9.0
3-03-76	4.7	0.72	15	0.71	15	0.55	12	6.0	3.3	59	11.8	59	732	672	5663	4876	9080	7890	108	8.8
3-04-76	-	0.76	-	0.71	15	0.55	12	-	1.6	58.5	11.7	58	948	776	4710	4150	7280	5950	104	8.8
3-05-76	4.7	0.75	16	0.71	15	0.55	12	6.0	3.4	58	11.6	58	680	656	5936	4710	7200	5840	100	9.0
3-06-76	-	-	-	-	-	-	-	-	2.6	63	-	58	2336	2286	4130	3570	7040	6080	108	9.0
3-07-76	-	0.75	16	0.70	15	-	-	-	1.9	68	-	58	814	692	4770	4140	7090	6110	114	9.1
AVERAGE	4.7	0.75	16	0.71	15	0.56	12	6.0	2.0	63.3	12.5	59	964	867	5391	4499	7725	6549	103.8	8.8

* SECONDARY CLARIFIER UNDERFLOW

TABLE A-5. PHASE II ORGANIC & SUSPENDED SOLIDS REMOVALS FOR PHOSTRIP PILOT PLANT - RENO/SPARKS, NEVADA

DATE	COD, mg/1		TSS, mg/1		VSS, mg/1	
	INF	EFF	INF	EFF	INF	EFF
2-23-76	237	13	115	27	-	21
2-24-76	238	64	99	-	92	-
2-25-76	226	27	97	19	95	14
2-26-76	238	27	101	15	97	11
2-27-76	226	27	111	33	105	27
2-28-76	206	20	112	32	106	30
2-29-76	241	25	123	33	92	32
3-01-76	204	20	-	3	-	2
3-02-76	216	20	43	-	32	-
3-03-76	241	20	77	1	77	1
3-04-76	238	42	101	13	76	9
3-05-76	234	17	59	4	51	3
3-06-76	277	24	72	9	60	8
3-07-76	243	24	92	15	79	11
AVERAGE	233	26	93	17	80	14

TABLE A-6. PHASE II PHOSPHORUS BALANCE DATA FOR PHOSTRIP PILOT PLANT - RENO/SPARKS, NEVADA

DATE	INFLUENT TOTAL P mg/l	INFLUENT ORTHO P mg/l	EFFLUENT TOTAL P mg/l	EFFLUENT ORTHO P mg/l	STRIPPER SUPERNATE TOTAL P mg/l	STRIPPER SUPERNATE ORTHO P mg/l	AERATION EFFLUENT TOTAL P FILTERED mg/l	AERATION EFFLUENT TOTAL P mg/l	* STRIPPER INFLUENT TOTAL P mg/l	WASTE SLUDGE TOTAL P mg/l	STRIPPER UNDERFLOW TOTAL P mg/l	STRIPPER UNDERFLOW FILTERED TOTAL P mg/l	CLARIFIER UNDERFLOW FILTERED TOTAL P mg/l	CHEMICAL CLARIFIER SUPERNATANT TOTAL P mg/l	CHEMICAL CLARIFIER SUPERNATANT ORTHO P mg/l
2-23-76	8.5	7.0	1.4	.78	36	36	1.4	32	195	195	230	25	3.0	-	1.2
2-24-76	9.2	7.8	1.17	1.12	40	37	1.9	33	210	210	225	27	2.2	-	-
2-25-76	8.7	7.8	.98	.79	37	35	.5	30	-	-	265	33	4.0	-	-
2-26-76	8.2	7.0	.54	.33	35	31	.5	32	200	200	225	33	1.5	-	2.6
2-27-76	8.5	7.2	.47	.26	35	35	.6	25	200	200	240	32	2.3	-	-
2-28-76	8.0	7.0	.46	.28	34	34	.6	28	170	170	210	28	.8	-	-
2-29-76	9.7	8.4	.72	.51	34	34	.8	30	195	195	185	36	1.4	-	-
3-01-76	8.0	6.2	.51	.33	33	31	.5	34	200	200	220	29	.9	-	5.6
3-02-76	7.3	5.4	.48	.21	29	27	.7	33	180	180	225	27	1.1	2.9	2.4
3-03-76	8.6	5.0	.55	.28	32	28	.6	31	185	185	220	32	.9	8.5	5.6
3-04-76	7.1	4.8	.56	.30	28	27	.7	20	175	175	185	27	1.4	3.5	5.2
3-05-76	8.0	5.6	.73	.43	27	25	.8	25	215	215	185	34	1.1	8.0	3.2
3-06-76	8.2	6.2	.82	.66	34	33	.5	29	175	175	190	29	1.0	7.5	2.6
3-07-76	9.4	6.2	.74	.51	33	30.5	.5	28	170	170	225	33	1.1	9.9	3.4
AVERAGE	8.4	6.5	.72	.49	33	32	.76	29	190	190	216	30	1.6	6.7	3.5
MAXIMUM	9.4	8.4	1.4	1.12	40	37	1.9	34	215	215	265	36	4.0	9.9	5.6
MINIMUM	7.1	4.8	.46	.21	27	25	0.5	20	170	170	185	25	.8	2.9	1.2

*SECONDARY CLARIFIER UNDERFLOW

TABLE A-7. PHASE III OPERATING CONDITIONS
FOR PHOSTRIP PILOT PLANT - RENO/SPARKS, NEVADA

DATE	PLANT INF GPM	ELUTRIATION		STRIPPER INFLUENT*		STRIPPER UNDERFLOW		AER RT_Q HRS	AVG D.O. mg/1	STRIPPER BLANKET HEIGHT INCHES	STRIPPER SRT HRS	M.L. TEMP °F	MLSS mg/1	MLVSS mg/1	STRIPPER INFLUENT*		STRIPPER UNDERFLOW		WASTE SLUDGE GAL/DAY	CHEMICAL CLARIFIER INFLUENT pH
		GPM	%Q	GPM	%Q	GPM	%Q								RSS mg/1	RVSS mg/1	RSS mg/1	RVSS mg/1		
3-20-76	-	0.47	-	0.65	-	0.55	-	-	1.9	53	10.6	62	1352	1136	8890	7520	11090	9370	56	8.7
3-21-76	-	0.47	-	0.65	-	0.55	-	-	1.7	54	10.8	62	1216	1044	8710	7390	10250	7850	34	8.7
3-22-76	5.14	0.47	10	0.65	14	0.55	12	5.45	1.5	56	11.2	62	1212	1044	9500	7990	10860	9190	30	11.2
3-23-76	5.14	0.47	10	0.65	14	0.55	12	5.45	1.2	58	11.6	62	1124	1028	8540	7270	10490	8840	54	7.7
3-24-76	5.14	-	-	0.65	14	0.55	12	5.45	1.3	55	11.0	63	1204	1008	8484	7144	12006	10145	52	8.7
3-25-76	5.14	0.47	10	0.65	14	0.55	12	5.45	1.5	56	11.2	62	1052	928	8030	6880	10020	8790	48	9.2
3-26-76	5.14	0.47	10	0.65	14	0.55	12	5.45	1.8	55	11.0	62	1124	1032	7850	6720	10400	9000	42	9.8
3-27-76	-	-	-	-	-	-	-	-	1.5	56	-	62	955	870	7910	6840	10660	9240	36	9.2
3-28-76	5.14	0.47	10	0.65	14	0.55	12	5.45	1.8	61	12.2	62	1096	1012	5670	6770	11220	9740	44	9.2
3-29-76	5.15	0.47	10	0.65	14	0.54	12	5.46	2.3	63	12.9	62	1180	1076	7550	6530	10170	8860	58	10.4
3-30-76	5.14	0.47	10	0.65	14	0.55	12	5.45	1.3	55	11.0	62	1000	985	8200	6980	10260	8750	42	9.5
3-31-76	5.15	0.47,	10	0.65	14	0.54	12	5.46	1.7	58	11.8	63	870	790	8230	6900	10260	8840	38	9.6
4-01-76	5.14	0.47	10	0.65	14	0.55	12	5.45	1.3	57	11.4	62	590	510	8200	6990	10360	8850	48	9.3
4-02-76	5.15	0.47	10	0.65	14	0.54	12	5.46	1.1	63	12.9	62	960	810	7180	6240	9560	8170	54	10.4
AVERAGE	5.14	0.47	10	0.65	14	0.55	12	5.45	1.6	57.1	11.4	62	1067	948	8067	7012	10543	8974	45.4	9.4

* SECONDARY CLARIFIER UNDERFLOW

TABLE A-8. PHASE III ORGANIC & SUSPENDED SOLIDS REMOVALS FOR PHOSTRIP PILOT PLANT - RENO/SPARKS, NEVADA

DATE	COD, mg/1		TSS, mg/1		VSS, mg/1	
	INF	EFF	INF	EFF	INF	EFF
3-20-76	301	27	102	20	98	13
3-21-76	256	32	97	20	94	17
3-22-76	223	40	107	27	105	21
3-23-76	223	-	120	42	115	27
3-24-76	239	27	117	33	112	32
3-25-76	261	33	89	39	87	25
3-26-76	261	37	107	40	106	26
3-27-76	253	41	77	13	67	10
3-28-76	212	33	89	22	-	15
3-29-76	233	33	100	15	-	15
3-30-76	207	24	64	11	64	9
3-31-76	219	32	95	2	95	1
4-01-76	235	12	116	39	85	33
4-02-76	247	10	87	54	84	49
AVERAGE	241	29	98	27	93	21

TABLE A-9. PHASE III PHOSPHORUS BALANCE DATA FOR PHOSTRIP PILOT PLANT - RENO/SPARKS, NEVADA

	INFLUENT		EFFLUENT		STRIPPER SUPERNATE		AERATION EFFLUENT		* STRIPPER INFLUENT	WASTE SLUDGE	STRIPPER UNDERFLOW		CLARIFIER UNDERFLOW FILTERED	CHEMICAL CLARIFIER SUPERNATANT	
DATE	TOTAL P mg/1	ORTHO P mg/1	TOTAL P mg/1	ORTHO P mg/1	TOTAL P mg/1	ORTHO P mg/1	TOTAL P FILTERED mg/1	TOTAL P mg/1	TOTAL P mg/1	TOTAL P mg/1	TOTAL P mg/1	FILTERED TOTAL P mg/1	TOTAL P mg/1	TOTAL P mg/1	ORTHO P mg/1
3-20-76	8.3	6.8	.79	.69	52	50	0.7	48	320	320	365	47	3.8	19.8	-
3-21-76	9.6	7.8	.72	.62	53	51.5	0.8	51	350	350	385	46	5.3	13.4	13.5
3-22-76	9.4	8.3	.77	.69	57	56	1.2	44	410	410	400	44	8.8	29.2	24.5
3-23-76	10.1	8.8	1.8	-	64	62.5	3.5	-	395	395	410	-	-	-	39
3-24-76	10.0	8.5	1.5	1.44	64	64	0.8	-	385	385	415	-	-	-	6.3
3-25-76	9.5	8.4	.5	.31	55	61	0.2	48	335	335	375	47	6.9	-	2.4
3-26-76	10.7	9.0	1.2	1.04	53	58	1.3	48	330	330	375	51	4.7	-	.4
3-27-76	10.5	8.5	1.1	0.6	60	65	0.6	43	300	300	355	52	4.4	12	4.2
3-28-76	10.3	8.7	.5	.26	57	61	0.4	-	355	355	385	-	-	-	3.8
3-29-76	10.5	8.5	.5	.25	57	61	0.5	-	350	350	340	-	-	-	.3
3-30-76	9.2	7.7	.4	.24	56	60	0.4	52	370	370	540	48	5.8	5.8	1.0
3-31-76	7.8	6.8	.4	.1	52	56.5	0.4	48	350	350	380	41	7.1	7.1	.4
4-01-76	7.7	6.4	.5	.18	55	53	0.5	48	185	185	345	38	2.0	2.0	1.6
4-02-76	7.6	6.4	.5	.14	51	51	0.4	45	300	300	340	38	0.9	0.9	.8
AVERAGE	9.4	7.9	.8	.61	56	58	.84	48	338	338	386	45	5.0	11.3	2.4**
MAXIMUM	10.7	9.0	1.8	1.44	64	65	3.5	52	410	410	540	52	8.8	29.2	39
MINIMUM	7.6	6.4	.4	.1	51	50	0.4	43	185	185	340	38	.9	.9	.3

* Secondary Clarifier Underflow

**MEDIAN

TABLE A-10. FULL SCALE PERFORMANCE DATA
PHOSTRIP SLUDGE RECYCLE SYSTEM
RENO/SPARKS, NEVADA
pH & TEMPERATURE °C*

DATE	FEED		EFF		AEROBIC RECYCLE		ANOXIC RECYCLE		STRIPPER SUPERNATANT		AERATION BAY END			AERATION RATE FT^3/GAL
	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX	TEMP	
9-13-75	7.1	7.4	7.2	7.7	6.7	6.8	6.4	6.4	6.5	6.5	-	7.0	24	1.5
9-14-75	7.1	7.2	7.3	7.6	6.7	6.6	6.4	6.3	6.5	6.4	-	6.7	23	1.5
9-15-75	7.0	7.3	7.2	7.6	6.7	6.6	6.3	6.3	6.4	6.4	-	-	23.5	1.6
9-16-75	7.2	7.5	7.3	7.7	6.6	6.8	6.4	6.4	6.6	6.5	-	7.0	23	1.6
9-17-75	7.1	7.4	7.5	7.7	6.8	6.7	6.4	6.3	6.5	6.4	-	-	23	1.6
9-18-75	7.1	7.4	7.1	7.6	6.8	6.9	6.4	6.5	6.6	6.6	-	7.0	23.5	1.6
9-19-75	7.1	7.3	7.2	7.4	6.8	6.9	6.4	6.5	6.5	6.5	-	-	23.5	1.6
9-20-75	7.0	7.3	7.1	7.4	6.8	6.7	6.4	6.2	6.5	6.4	-	6.7	23.5	1.6
9-21-75	6.9	6.9	7.0	7.3	6.7	6.7	6.4	6.3	6.5	6.5	-	-	23	1.5
9-22-75	6.9	7.3	7.1	7.5	6.8	6.8	6.4	6.4	6.6	6.5	-	6.9	23	1.5
9-23-75	7.1	7.2	7.2	7.5	6.8	6.7	6.4	6.4	6.7	6.5	-	-	23	1.5
9-24-75	6.9	7.7	7.2	7.6	6.9	6.9	6.4	6.4	6.6	6.6	-	7.2	23	1.3
9-25-75	6.9	7.8	7.2	-	6.8	6.8	6.5	6.4	6.7	6.5	-	-	23	1.2
9-26-75	6.9	-	6.9	-	6.8	6.8	6.5	6.4	6.6	6.6	-	6.8	23	1.3
9-27-75	7.0	7.2	7.1	7.4	6.8	6.8	6.4	6.4	6.6	6.5	-	-	23	1.6
MEDIAN	7.0	7.3	7.2	7.6	6.8	6.8	6.4	6.4	6.6	6.5	-	7.0	23	1.5

*GRAB SAMPLES

TABLE A-11. FULL SCALE PERFORMANCE DATA
PHOSTRIP SLUDGE RECYCLE SYSTEM
RENO/SPARKS, NEVADA
FLOW RATES (MGD)

DATE	FEED MIN	FEED MAX	AEROBIC RECYCLE MIN	AEROBIC RECYCLE MAX	ANOXIC RECYCLE MIN	ANOXIC RECYCLE MAX	STRIPPER SUPERNATANT MIN	STRIPPER SUPERNATANT MAX	WASTE (CLARIFIER)	SLUDGE RECYCLE MIN	SLUDGE RECYCLE MAX	BLANKET DEPTH (m) CLAR MIN/MAX	BLANKET DEPTH (m) STRIPPER MIN/MAX
9-13-75	4.9	7.0	1.37	1.78	.62	1.1	.75	.68	.08	1.12	.64	.1/.3	1.3/.9
9-14-75	4.6	5.9	1.17	1.65	.62	1.1	.55	.55	.178	1.12	.64	.05/.05	1.2/1.4
9-15-75	4.4	6.9	1.17	1.65	.62	1.1	.55	.55	.116	1.12	.64	.08/.08	1.2/1.3
9-16-75	3.8	6.7	1.17	1.94	.62	1.1	.55	.84	.134	1.12	.64	.05/.08	1.1/.9
9-17-75	3.8	6.8	1.17	1.80	.62	1.1	.55	.70	.128	1.12	.64	.08/.08	1.1/.9
9-18-75	3.6	6.9	1.17	1.91	.62	1.1	.55	.81	.153	1.12	.64	.05/.08	1.3/.9
9-19-75	4.0	6.8	1.5	1.98	.62	1.1	.88	.88	.259	1.12	.64	.05/.05	.9/.7
9-20-75	4.2	6.5	1.5	1.98	.62	1.1	.88	.88	.311	1.12	.64	.08/.05	1.0/1.0
9-21-75	4.0	6.2	1.5	1.98	.62	1.1	.88	.88	.318	1.12	.64	.08/.1	1.2/1.4
9-22-75	3.7	6.8	1.5	1.98	.62	1.1	.88	.88	.131	1.12	.64	.08/.1	1.4/1.3
9-23-75	3.7	6.8	1.5	1.98	.62	1.1	.88	.88	.073	1.12	.64	.1/.1	1.4/1.3
9-24-75	4.0	6.8	1.5	1.98	.62	1.1	.88	.88	.132	1.12	.64	.08/.08	1.1/1.1
9-25-75	3.9	5.8	1.5	1.98	.62	1.1	.88	.88	.215	1.12	.64	.1/.1	1.1/1.3
9-26-75	4.0	6.6	1.5	1.98	.62	1.1	.88	.88	.106	1.12	.64	.08/.05	1.6/1.5
9-27-75	3.9	-	1.5	1.98	.62	1.1	.88	.88	.107	1.12	.64	.08/.05	1.2/1.3
MEAN	3.9	6.6	1.38	1.94	.62	1.1	.76	.84	.17	1.12	.64	.08/.09	1.2/1.2

TABLE A-12. FULL SCALE PERFORMANCE DATA
PHOSTRIP SLUDGE RECYCLE SYSTEM
RENO/SPARKS, NEVADA
ORTHO PHOSPHORUS (mg/1)

DATE	FEED		EFFLUENT		AEROBIC RECYCLE		ANOXIC RECYCLE		STRIPPER SUPERNATANT		AERATION GRAB	
	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX
9-13-75	5.0	6.5	.15	.85	11	10	64	52	48	44	-	.02
9-14-75	6.2	7.6	-	.35	14	12.5	64	55	43	45	-	.05
9-15-75	6.8	7.4	.55	.6	13	12.5	69.5	60	42	43	-	-
9-16-75	6.0	5.5	.25	.23	18	14.5	83	74	55	51	-	.11
9-17-75	5.1	6.2	.15	.15	14	7.5	74	72	48	50	-	-
9-18-75	6.4	7.2	.28	.25	5.5	13.5	76	68	46	49	-	.02
9-19-75	6.2	6.8	.08	.10	2.0	7.5	72	55	47	45	-	-
9-20-75	5.8	7.2	.08	.25	3.8	4.0	60	60	40	42	-	.03
9-21-75	5.8	7.6	.05	.13	2	.05	56	56	36	36	-	-
9-22-75	6.2	6.8	.15	.33	5.3	6.5	60	43	32	31	-	.01
9-23-75	5.7	6.3	.18	.55	3.5	7.0	57	50	36	33	-	-
9-24-75	19.0	5.7	.98	.20	6.5	9.0	68	63	45	44	-	.04
9-25-75	16.5	6.0	.98	-	6.5	11.5	68	62	39	39	-	-
9-26-75	6.0	5.7	.48	1.0	4.5	-	52	46	33	28	-	.02
9-27-75	5.3	7.0	.38	.3	4	2	66	58	43	40	-	-
MEAN	7.5	6.6	.33	.38	7.6	8.4	66	58	42	41	-	.04
MINIMUM	5.0	5.5	.05	.1	2	.05	52	43	32	28	-	.01
MAXIMUM	19.0	7.6	.98	1.0	18	14.5	83	72	55	51	-	.11

TABLE A-13. FULL SCALE PERFORMANCE DATA
PHOSTRIP SLUDGE RECYCLE SYSTEM
RENO/SPARKS, NEVADA
SUSPENDED SOLIDS (mg/l)

	INFLUENT		EFFLUENT		MIXED LIQUOR			ANOXIC RECYCLE			AEROBIC RECYCLE			STRIPPER SUPERNATANT		
	TSS	VSS/TSS	TSS	VSS/TSS	TSS		VSS/TSS	TSS		VSS/TSS	TSS		VSS/TSS	TSS		VSS/TSS
DATE	C		MIN/MAX		MIN	MAX		MIN	MAX		MIN	MAX		MIN	MAX	
9-13-75	132	-	20C	-	1140C	-	.79	-	-	-	5237	8158C	.79	-	164	.51
9-14-75	96	-	11C	-	1304C	1200	.78	10896	10144	.77	7572C	4916	.78	8	116	.50
9-15-75	113	-	19C	-	1122C	629	.75	10376	10172	.77	3088	4860	.81	20	-	.69
9-16-75	198	-	14C	-	1154C	673	.73	12396	9692	.78	8608	6360	.79	60	92	.84
9-17-75	113	-	15C	-	1296C	716	.76	-	10700	.79	1848	6292	.80	88	112	.72
9-18-75	105	-	10C	-	1150C	1320	.82	11972	10140	.79	1792	5864	.79	84	132	.89
9-19-75	122	-	12C	-	1078C	460	-	10272	8404	.80	1756	5032	.78	64	96	.76
9-20-75	76	-	9C	-	938C	507	.83	9672	8752	.81	2560	3952	.83	110	99	-
9-21-75	79	-	8C	-	974C	-	-	8972	7292	.81	2344	2784	.84	110	67	.9
9-22-75	118	-	27/15	.86	1068C	-	-	6400	7992	.84	2532	3444	.81	58	36	.84
9-23-75	106	-	10/18	.6	1168	680	.82	8632	7980	.82	-	5316	.83	35	73	.83
9-24-75	113	-	23/12	.66	992C	-	-	7688	8564	.78	2860	4016	.81	68	72	.82
9-25-75	93	-	31/-	.87	1178C	612	.78	8804	8748	.82	2684	4428	.83	54	60	.83
9-26-75	133	-	15C	-	1306C	-	-	10528	8236	.83	2316	4332	.83	85	52	.87
9-27-75	104	-	6/7	-	1356C	-	-	8508	10924	.83	2168	3064	.77	-	87	-
MEAN	-	-	19/13	.75	1168	755	.78	9624	9124	.80	3061	4618	.81	65	90	.77
MINIMUM	76	-	6/7	-	1168	460	.73	6400	7292	.77	1756	2784	.77	8	36	.50
MAXIMUM	198	-	31/18	-	1168	1320	.83	12396	10924	.84	8608	6360	.84	110	164	.90
MEAN C	113	-	13	-	1147	-	.78	-	-	-	7572	8158	.79	-	-	-

C = 24 HR COMPOSITE

TABLE A-14. FULL SCALE PERFORMANCE DATA
PHOSTRIP SLUDGE RECYCLE SYSTEM
RENO/SPARKS, NEVADA
BOD_5, COD, DISSOLVED OXYGEN (mg/1)

							DISSOLVED OXYGEN									
	BOD_5		COD						AEROBIC RECYCLE		ANOXIC RECYCLE		STRIPPER SUPERNATANT		END OF AERATION	
			INF		EFF		EFF									
DATE	INF	EFF	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX
9-13-75	-	-	196	243	48	44	-	-	-	-	-	-	-	-	-	-
9-14-75	-	-	272	264	40	48	-	2.6	-	.2	-	.3	-	.5	-	.7
9-15-75	-	-	188	285	37	36	-	-	-	-	-	-	-	-	-	-
9-16-75	-	-	188	247	44	44	-	2.9	-	.6	-	.2	-	.6	-	1.5
9-17-75	103	16	145	272	36	36	-	-	-	-	-	-	-	-	-	-
9-18-75	-	-	205	255	44	44	-	2.9	-	.4	-	.3	-	.8	-	1.8
9-19-75	-	-	206	243	35	31	-	-	-	-	-	-	-	-	-	-
9-20-75	-	-	156	256	39	35	-	3.3	-	.5	-	.4	-	.7	-	1.1
9-21-75	-	-	156	306	39	35	-	-	-	-	-	-	-	-	-	-
9-22-75	-	-	-	-	48	34	3.7	-	.4	-	.3	-	0.8	-	-	1.8
9-23-75	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
9-24-75	-	-	181	276	36	32	-	4.9	-	1.6	-	0.2	-	1.5	-	2.2
9-25-75	-	-	151	260	39	34	-	-	-	-	-	-	-	-	-	-
9-26-75	-	-	194	288	36	33	-	-	-	-	-	-	-	-	-	-
9-27-75	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MEAN	103	16	187	266	40	37	3.7	3.3	.4	.7	.3	.3	.8	.8	-	1.5

TABLE A-15. FULL SCALE PERFORMANCE DATA
PHOSTRIP SLUDGE RECYCLE SYSTEM
RENO/SPARKS, NEVADA
TOTAL PHOSPHORUS (mg/1)

DATE	FEED		EFFLUENT		AERATION		AEROBIC RECYCLE				ANOXIC RECYCLE				STRIPPER SUPERNATANT	
	MIN	MAX	MIN	MAX	MIN	MAX	MIN	MAX	MIN FILT	MAX FILT	MIN	MAX	MIN FILT	MAX FILT	MIN	MAX
9-13-75	7.4	8.8	1.0	-	-	-	123	-	-	-	213	-	-	-	48.5	-
9-14-75	9.0	9.6	-	1.5	-	33	103	138	-	-	-	206	70.5	60	48.5	45
9-15-75	8.4	9.5	1.3	1.4	-	29	114	198	-	-	250	244	73.5	62	45	46.5
9-16-75	7.8	7.8	1.2	1.2	-	34	-	178	-	-	260	256	78	71	49.5	48
9-17-75	6.4	8.5	1.0	0.8	-	-	122	206	-	-	276	286	77	67	46	46.5
9-18-75	8.2	8.9	1.2	1.2	-	30	62	184	-	-	294	266	79	71	49.5	49.5
9-19-75	8.2	9.1	0.5	0.5	-	26	80	176	-	-	266	240	75	58.5	50	53
9-20-75	8.3	9.3	0.5	0.6	-	28	104	150	8	6	252	220	62.5	-	45.5	41.5
9-21-75	7.4	10.2	0.6	0.5	-	-	98	114	5.5	4	250	222	54.5	49.5	38	36.5
9-22-75	8.4	9.9	0.7	0.7	27	27	110	154	-	-	264	266	63	50	36	34
9-23-75	7.1	9.3	0.9	0.8	-	-	126	190	6	8	300	248	56	51	37	37
9-24-75	21	8.5	1.6	0.9	-	27	118	156	7.5	9	262	244	68	62	44	42
9-25-75	18.4	8.9	1.4	-	-	-	110	150	8	13.5	238	208	66	62	44	43.5
9-26-75	9.5	8.5	0.9	1.0	-	36	76	140	6	7.5	244	206	53	48	35	32
9-27-75	8.1	9.6	0.8	0.4	-	-	108	116	4	4.5	228	248	-	61	-	24
MEAN	9.6	9.1	1.0	0.8	27	30	104	161	6.4	7.5	257	240	67	55	44	41
MINIMUM	6.4	7.8	.5	0.4	27	26	62	116	4	4	213	206	53	48	35	24
MAXIMUM	21	10.2	1.6	1.5	27	36	126	206	8	13.5	300	286	79	71	50	53

TABLE A-16. FULL SCALE PERFORMANCE DATA
PHOSTRIP SLUDGE RECYCLE SYSTEM
RENO/SPARKS, NEVADA
(CONTROL)

DATE	FEED FLOW (MGD)		RECYCLE (MGD) AVE	MLSS mg/1 AVE	RSS mg/1 AVE	AERATION RATE FT^3/GAL	ESS mg/1	WAS MGD
	MIN	MAX						
9-13-75	4.9	7.0	1.91	1,046	4,742	1.1	10	.204
9-14-75	4.6	5.9	1.97	974	4,986	1.1	9	.204
9-15-75	4.4	6.9	2.01	1,048	4,274	1.4	20	.219
9-16-75	3.8	6.7	1.98	1,078	3,686	1.5	8	.205
9-17-75	3.8	6.8	1.96	1,200	4,270	1.4	15	.204
9-18-75	3.6	6.9	1.80	1,014	3,290	1.5	18	.194
9-19-75	4.0	6.8	1.66	1,050	4,248	1.4	14	.186
9-20-75	4.2	6.5	1.50	940	5,242	1.4	7	.162
9-21-75	4.0	6.2	1.53	996	4,362	1.4	9	.168
9-22-75	3.7	6.8	1.80	924	4,650	1.4	19	.195
9-23-75	3.7	6.8	2.47	894	3,204	1.4	12	.253
9-24-75	4.0	6.8	1.97	868	3,366	1.4	14	.210
9-25-75	3.9	5.8	1.82	886	3,728	1.3	15	.198
9-26-75	4.0	6.6	1.76	1,050	3,440	1.2	22	.197
9-27-75	3.9	-	1.63	1,014	3,410	1.5	18	.175
MEAN	3.9	6.6	1.85	1,000	4,060	1.4	14	.198

TABLE A-17. FULL SCALE PERFORMANCE DATA
PHOSTRIP LPE SYSTEM - RENO/SPARKS, NEVADA
FLOW RATES, MGD

			AEROBIC RECYCLE				ANOXIC RECYCLE				STRIPPER SUPERNATANT		STRIPPER ELUTRIANT	TOTAL RETURN		WASTE	
			1		2		1		2								
DATE	FEED	FEED HIGH/LOW	HIGH	LOW	HIGH	LOW	HIGH	LOW	HIGH	LOW	HIGH	LOW		1	2	1	2
11-22-76	11.44	-	-	-	-	-	0.52	-	0.58	-	1.68	-	1.10	-	-	-	-
11-23-76	11.89	-	-	-	-	-	0.65	-	0.60	-	1.69	-	1.15	-	-	-	-
11-24-76	12.00	-	-	-	-	-	0.63	-	0.60	-	1.61	-	1.03	-	-	-	-
11-25-76	10.98	-	-	-	-	-	0.60	-	0.60	-	1.53	-	0.98	-	-	-	-
11-26-76	12.00	-	-	-	-	-	0.52	-	0.63	-	1.65	-	-	-	-	-	-
11-27-76	11.80	-	-	-	-	-	0.35	-	0.56	-	1.45	-	0.76	-	-	-	-
11-28-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-29-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-30-76	-	-	-	-	-	-	0.71	-	0.61	-	1.69	-	1.18	-	-	-	-
12-01-76	11.69	-	-	-	-	-	-	-	-	-	-	-	-	1.58	1.70	.121	.118
12-02-76	11.87	-	-	-	-	-	-	-	-	-	-	-	-	1.45	1.46	.085	.103
12-03-76	11.78	-	-	-	-	-	-	-	-	-	-	-	-	1.51	1.75	.066	.097
12-04-76	11.82	-	-	-	-	-	-	-	-	-	-	-	-	1.66	1.67	.093	.094
12-05-76	11.61	-	-	-	-	-	-	-	-	-	-	-	-	1.69	1.82	.116	.105
12-06-76	11.80	-	1.56	1.56	1.31	1.31	0.76	0.76	0.59	0.59	1.96	-	0.50	1.53	1.60	.090	.093
12-07-76	11.47	-	1.26	1.26	1.16	1.16	0.77	0.77	0.70	0.70	2.24	-	1.29	1.37	1.53	.071	.090
12-08-76	11.49	-	-	-	-	-	-	-	-	-	-	-	-	1.45	1.36	.070	.078
12-09-76	11.57	-	1.40	1.40	1.21	1.21	0.79	0.79	0.84	0.84	1.56	-	0.64	1.48	1.36	.071	.081
12-10-76	11.52	-	1.35	1.35	1.15	1.15	0.78	0.78	0.84	0.84	1.96	-	1.14	1.49	1.33	.073	.078
12-11-76	11.54	-	-	-	-	-	-	-	-	-	-	-	-	1.50	1.28	.084	.090
12-12-76	11.14	-	-	-	-	-	-	-	-	-	-	-	-	1.45	1.34	.085	.093
12-13-76	11.53	-	-	-	-	-	-	-	-	-	-	-	-	1.50	1.33	.088	.090
12-14-76	11.48	-	-	-	-	-	-	-	-	-	-	-	-	1.48	1.32	.085	.093
12-15-76	11.15	-	1.33	1.33	1.07	1.07	0.78	0.78	0.83	0.83	1.49	-	1.40	1.43	1.39	.085	.093
12-16-76	11.15	-	1.34	1.34	1.05	1.05	0.73	0.73	0.77	0.77	1.61	-	1.28	1.43	1.29	.074	.089
12-17-76	11.15	-	1.35	1.35	0.95	0.95	0.65	0.65	0.76	0.76	1.87	-	1.15	1.40	1.42	.079	.092
12-18-76	11.40	-	-	-	-	-	-	-	-	-	-	-	-	1.42	1.44	.088	.076
12-19-76	11.20	-	-	-	-	-	-	-	-	-	-	-	-	1.48	1.47	.084	.103
12-20-76	11.80	-	-	-	-	-	-	-	-	-	-	-	-	1.35	1.45	.076	.096
12-21-76	11.60	-	-	-	-	-	-	-	-	-	-	-	-	1.31	1.45	.073	.101
12-22-76	11.80	-	-	-	-	-	-	-	-	-	-	-	-	1.33	1.41	.076	.095
12-23-76	11.80	-	-	-	-	-	-	-	-	-	-	-	-	1.45	1.57	.081	.104
12-24-76	12.20	-	-	-	-	-	-	-	-	-	-	-	-	1.60	1.53	.089	.135
12-25-76	10.60	-	-	-	-	-	-	-	-	-	-	-	-	1.41	1.44	.082	.101
12-26-76	11.40	-	-	-	-	-	-	-	-	-	-	-	-	1.42	1.43	.108	.102
12-27-76	12.20	-	-	-	-	-	-	-	-	-	-	-	-	1.41	1.43	.105	.097
12-28-76	12.40	-	-	-	-	-	-	-	-	-	-	-	-	1.45	1.61	.107	.105
12-29-76	12.40	-	0.59	-	0.91	-	0.62	-	0.62	-	1.73	-	1.66	1.44	1.63	.104	.108
12-30-76	12.40	-	-	0.58	-	0.59	-	0.07	-	0.33	2.09	-	1.66	1.54	1.69	.112	.114
12-31-76	12.60	-	-	-	-	-	-	-	-	-	-	-	-	1.49	1.55	.109	.103
01-01-77	13.20	-	-	-	-	-	-	-	-	-	-	-	HIGH/LOW	1.45	1.79	-	-
01-02-77	14.00	-	-	-	-	-	-	-	-	-	-	-	-	1.41	1.81	-	-
01-03-77	12.00	-	-	0.47	-	0.59	-	0.07	-	0.27	-	2.08	-/1.36	1.43	1.88	-	-
01-04-77	12.80	-	-	0.50	-	0.60	-	0.12	-	0.28	-	2.16	-/1.20	1.25	1.60	-	-
01-05-77	12.00	-	0.82	-	0.86	-	0.54	-	0.63	-	-	1.58	.58/-	1.43	1.94	-	-
01-06-77	12.20	15.3/6.7	0.89	0.88	0.92	1.24	0.59	0.48	0.63	0.36	2.12	1.51	.65/.65	1.34	1.87	-	-
01-07-77	11.20	14/9	0.86	0.63	0.91	0.53	0.57	0.32	0.65	0.33	1.37	1.51	.50/.43	1.17	1.74	-	-
01-08-77	12.00	-	-	-	-	-	-	-	-	-	-	-	-	1.05	1.87	-	-
01-09-77	11.40	-	-	-	-	-	-	-	-	-	-	-	-	1.04	1.55	-	-
01-10-77	12.80	15.3/7.3	0.89	0.52	0.91	0.53	0.51	0.29	0.60	0.35	1.73	1.51	.43/.36	1.17	1.68	-	-
01-11-77	12.80	15.3/7.3	0.89	0.49	0.91	1.51	0.66	0.32	0.69	0.35	2.45	2.02	1.51/.58	1.33	1.89	-	-
01-12-77	11.60	14/7.3	0.86	-	0.92	-	0.71	-	0.66	-	1.51	-	.79/-	1.21	1.76	-	-
01-13-77	11.60	14/7.3	0.88	0.48	0.92	0.53	0.63	0.27	0.69	0.39	1.44	1.37	.79/1.01	1.16	1.73	-	-
01-14-77	12.80	14/7.3	0.89	0.50	0.95	0.60	0.60	0.33	0.60	0.39	1.87	1.51	1.15/1.01	1.25	1.65	-	-
01-15-77	12.80	-	-	-	-	-	-	-	-	-	-	-	-	1.21	1.64	-	-
01-16-77	11.20	-	-	-	-	-	-	-	-	-	-	-	-	1.26	1.51	-	-
01-17-77	12.20	15.3/9	0.86	0.52	0.89	0.56	0.63	0.33	0.59	0.40	1.66	1.51	1.08/1.22	1.30	1.54	-	-
01-18-77	11.60	15.3/9	0.84	-	0.94	-	0.59	-	0.60	1.31	1.80	-	1.58/-	1.25	1.48	-	-
01-19-77	11.40	15.3/9	-	0.53	-	0.53	-	0.36	-	0.43	-	1.58	-/1.30	1.22	1.42	-	-
01-20-77	12.00	-	-	-	-	-	-	-	-	-	-	-	-	1.30	1.22	-	-
01-21-77	12.60	15.3/9	-	0.55	-	0.53	-	0.35	-	0.39	-	1.66	-/1.37	0.99	1.24	-	-
01-22-77	12.00	-	-	-	-	-	-	-	-	-	-	-	-	1.83	1.26	-	-
01-23-77	11.20	-	-	-	-	-	-	-	-	-	-	-	-	1.02	1.48	-	-
01-24-77	11.60	15.3/9	0.80	-	0.91	-	0.55	-	0.58	-	2.09	-	1.58/-	1.12	1.49	-	-
01-25-77	11.60	16.7/10	0.82	0.51	0.87	0.53	0.60	0.28	0.64	0.38	1.80	1.92	1.34/1.44	1.06	1.37	-	-
01-26-77	12.20	16/10	0.86	-	0.88	-	0.62	-	0.59	-	1.73	-	1.44/-	1.04	1.52	-	-
01-27-77	12.00	16/9.3	-	0.52	-	0.59	-	0.32	-	0.37	-	1.80	-/1.51	0.99	1.56	-	-
01-28-77	11.10	16.7/10	-	0.51	-	0.62	-	0.32	-	0.38	-	1.77	-/1.37	1.10	1.61	-	-
01-29-77	11.44	-	-	-	-	-	-	-	-	-	-	-	-	1.27	1.64	-	-
01-30-77	11.09	-	-	-	-	-	-	-	-	-	-	-	-	1.24	1.60	-	-

TABLE A-18. FULL SCALE PERFORMANCE DATA
PHOSTRIP LPE SYSTEM – RENO/SPARKS NEVADA
PH

FEED	FEED PH	EFFLUENT PH 1	PH 2	MIXED LIQUOR 1 HIGH	LOW	MIXED LIQUOR 2 HIGH	LOW	STRIPPER SUPERNATANT
11-22-76	-	-	-	-	-	-	-	-
11-23-76	-	-	-	7.3	-	7.3	-	-
11-24-76	-	8.3	8.4	8.0	7.3	7.9	7.2	7.8
11-25-76	8.2	8.3	8.6	7.6	7.8	7.6	7.8	-
11-26-76	7.8	8.2	7.9	7.6	7.4	7.6	7.4	7.7
12-27-76	8.0	7.7	7.8	7.4	8.2	7.4	7.8	8.1
11-28-76	-	-	-	-	-	-	-	-
11-29-76	-	-	-	-	-	-	-	-
11-30-76	7.7	8.4	7.8	7.3	7.8	7.3	7.6	7.8
11-31-76	-	-	-	-	-	-	-	-
12-01-76	7.7	7.8	7.8	7.3	7.1	7.3	7.2	7.6
12-02-76	7.8	7.7	7.7	-	7.3	-	7.3	7.3
12-03-76	7.5	7.5	7.5	-	7.0	-	7.0	-
12-04-76	-	-	-	-	-	-	-	-
12-05-76	-	-	-	-	-	-	-	-
12-06-76	7.5	7.5	7.5	7.5	7.3	7.6	7.3	7.5
12-07-76	8.4	8.1	8.3	-	-	-	-	8.3
12-08-76	8.4	8.1	8.3	7.6	7.2	7.7	7.5	8.2
12-09-76	8.8	8.6	8.7	-	-	-	-	8.4
12-10-76	8.8	8.2	8.2	7.2	7.4	7.3	7.6	8.8
12-11-76	-	-	-	-	-	-	-	-
12-12-76	-	-	-	-	-	-	-	-
12-13-76	-	-	-	-	-	-	-	-
12-14-76	-	-	-	-	-	-	-	-
12-15-76	8.3	8.3	7.9	7.6	7.4	7.5	7.3	7.7
12-16-76	7.5	7.6	7.6	-	-	-	-	7.5
12-17-76	7.7	7.8	7.8	-	-	-	-	7.6
12-18-76	-	-	-	-	-	-	-	-
12-19-76	-	-	-	-	-	-	-	-
12-20-76	-	-	-	-	-	-	-	-
12-21-76	-	-	-	-	-	-	-	-
12-22-76	-	-	-	-	-	-	-	-
12-23-76	-	-	-	-	-	-	-	-
12-24-76	-	-	-	-	-	-	-	-
12-25-76	-	-	-	-	-	-	-	-
12-26-76	-	-	-	-	-	-	-	-
12-27-76	7.2	7.4	7.4	-	-	-	-	-
12-28-76	7.5	7.3	7.3	7.0	7.1	7.0	7.0	7.2
12-29-76	7.3	7.1	7.3	6.9	6.9	6.8	6.9	6.9
12-30-76	-	-	-	-	-	-	-	-
12-31-76	-	-	-	-	-	-	-	-
01-01-77	-	-	-	-	-	-	-	-
01-02-77	7.1	7.1	7.2	-	-	-	-	6.8
01-03-77	7.7	7.5	7.7	6.9	6.7	7.0	6.8	7.2
01-04-77	-	7.3	7.4	-	-	-	-	7.0
01-05-77	7.4	7.5	7.6	7.2	6.8	7.2	6.9	-
01-06-77	7.7	7.7	7.9	-	-	-	-	7.4
01-07-77	-	-	-	7.1	7.0	7.2	7.1	-
01-08-77	-	-	-	-	-	-	-	-
01-09-77	7.3	7.3	7.5	-	-	-	-	-
01-10-77	7.6	7.4	7.6	7.0	6.9	7.1	7.0	7.1
01-11-77	7.5	7.5	7.5	-	-	-	-	7.0
01-12-77	7.4	7.5	7.6	7.0	6.9	7.0	6.9	7.1
01-13-77	7.7	7.3	7.5	-	-	-	-	7.0
01-14-77	-	-	-	7.0	6.9	7.0	6.9	-
01-15-77	-	-	-	-	-	-	-	-
01-16-77	7.3	7.3	7.5	7.1	7.1	7.0	7.0	7.0
01-17-77	7.4	7.7	7.6	-	-	-	-	7.1
01-18-77	7.5	7.2	7.5	-	-	-	-	7.3
01-19-77	7.8	7.3	7.4	7.0	7.0	7.0	6.9	7.3
01-20-77	7.5	7.3	7.5	-	-	-	-	-
01-21-77	-	-	-	-	-	-	-	-
01-22-77	-	-	-	-	-	-	-	-
01-23-77	7.6	7.3	7.4	-	-	-	-	7.4
01-24-77	7.7	7.6	7.7	6.9	7.2	6.8	7.0	7.4
01-25-77	7.8	7.2	7.4	-	-	-	-	7.2
01-26-77	7.8	7.8	7.8	7.2	6.9	7.2	6.9	7.6
01-27-77	8.2	7.9	8.0	-	-	-	-	8.3
01-28-77	-	-	-	7.0	7.2	7.0	7.2	-
01-29-77	-	-	-	-	-	-	-	-
01-30-77	8.0	7.7	7.8	-	-	-	-	7.5

TABLE A-19. FULL SCALE PERFORMANCE DATA
PHOSTRIP LPE SYSTEM - RENO/SPARKS, NEVADA
BOD$_5$, COD, D.O. (mg/l)

	BOD$_5$ INFLUENT		BOD$_5$ EFFLUENT		COD INFLUENT	D.O. STRIPPER SUPERNATANT
DATE	1	2	1	2		
11-22-76	-	-	-	-	-	-
11-23-76	-	-	-	-	-	-
11-24-76	-	-	-	-	-	-
11-25-76	-	-	-	-	-	-
11-26-76	-	-	-	-	-	-
11-27-76	-	-	-	-	-	-
11-28-76	-	-	-	-	-	-
11-29-76	-	-	-	-	-	-
11-30-76	-	-	-	-	-	-
11-31-76	-	-	-	-	-	-
12-01-76	-	-	-	-	-	-
12-02-76	133	163	17	25	-	-
12-03-76	-	-	-	-	-	-
12-04-76	-	-	-	-	-	-
12-05-76	-	-	-	-	-	-
12-06-76	158	-	-	-	-	-
12-07-76	133	-	-	-	-	-
12-08-76	-	-	-	-	-	-
12-09-76	163	165	23	22	-	-
12-10-76	-	-	-	-	-	-
12-11-76	-	-	-	-	-	-
12-12-76	-	-	-	-	-	-
12-13-76	-	-	-	-	-	-
12-14-76	-	-	-	-	-	-
12-15-76	-	-	-	-	-	-
12-16-76	147	130	29	28	-	2.4
12-17-76	-	-	-	-	-	-
12-18-76	-	-	-	-	-	-
12-19-76	-	-	-	-	-	-
12-20-76	-	-	-	-	-	-
12-21-76	-	-	-	-	-	-
12-22-76	153	155	29	30	-	-
12-23-76	-	-	-	-	-	-
12-24-76	-	-	-	-	-	-
12-25-76	-	-	-	-	-	-
12-26-76	-	-	-	-	-	-
12-27-76	-	-	-	-	-	-
12-28-76	-	140	-	28	-	3.1
12-29-76	143	160	-	-	-	-
12-30-76	-	-	-	-	-	-
12-31-76	-	-	-	-	-	-
01-01-77	-	-	-	-	-	-
01-02-77	-	-	-	-	-	-
01-03-77	-	-	-	-	-	-
01-04-77	-	-	-	-	-	-
01-05-77	-	-	-	-	-	-
01-06-77	133	133	-	20	-	-
01-07-77	-	-	-	-	-	-
01-08-77	-	-	-	-	-	-
01-09-77	-	-	-	-	-	-
01-10-77	-	80	22	20	-	-
01-11-77	-	-	-	-	-	-
01-12-77	-	-	-	-	-	-
01-13-77	143	163	24	25	-	-
01-14-77	-	-	-	-	-	-
01-15-77	-	-	-	-	-	-
01-16-77	-	-	-	-	-	-
01-17-77	-	-	-	-	-	-
01-18-77	-	201	34	34	-	-
01-19-77	-	-	-	-	-	-
01-20-77	-	177	25	26	-	-
01-21-77	-	-	-	-	-	-
01-22-77	-	-	-	-	-	-
01-23-77	160	193	29	31	-	-
01-24-77	-	-	-	-	-	-
01-25-77	-	163	27	-	-	-
01-26-77	-	-	-	-	-	-
01-27-77	135	145	25	25	-	-
01-28-77	-	-	-	-	-	-
01-29-77	-	-	-	-	-	-
01-30-77	-	-	-	-	-	-

TABLE A-20. FULL SCALE PERFORMANCE DATA
PHOSTRIP LPE SYSTEM - RENO/SPARKS, NEVADA
TOTAL-PHOSPHORUS (mg/1)

DATE	FEED	EFFLUENT 1	EFFLUENT 2	AEROBIC RECYCLE 1 HIGH	AEROBIC RECYCLE 1 LOW	AEROBIC RECYCLE 2 HIGH	AEROBIC RECYCLE 2 LOW	ANOXIC RECYCLE HIGH	ANOXIC RECYCLE LOW	STRIPPER SUPERNATANT
11-22-76	-	-	-	-	-	-	-	-	-	-
11-23-76	-	-	-	-	-	-	-	-	-	-
11-24-76	-	-	-	-	-	-	-	-	-	-
11-25-76	-	-	-	-	-	-	-	-	-	-
11-26-76	-	-	-	168	88	-	120	232	176	-
11-27-76	-	-	-	-	-	-	-	-	-	-
11-28-76	-	-	-	-	-	-	-	-	-	-
11-29-76	-	-	-	-	-	-	-	-	-	-
11-30-76	-	-	-	-	-	-	-	-	-	-
11-31-76	-	-	-	-	-	-	-	-	-	-
12-01-76	-	-	-	208	80	168	112	168	168	-
12-02-76	-	-	-	-	-	-	-	-	-	-
12-03-76	-	-	-	-	-	-	-	-	-	-
12-04-76	-	-	-	-	-	-	-	-	-	-
12-05-76	-	-	-	-	-	-	-	-	-	-
12-06-76	10	3.3	3.5	184	120	116	88	160	176	19
12-07-76	11.2	6.1	7.2	-	-	-	-	-	-	43
12-08-76	6.6	3.1	2.6	-	-	-	-	-	-	27
12-09-76	9.4	3.3	3.9	-	-	-	-	-	-	19
12-10-76	8.4	2.5	3.7	-	-	-	-	-	-	23
12-11-76	-	-	-	-	-	-	-	-	-	-
12-12-76	-	-	-	-	-	-	-	-	-	-
12-13-76	-	-	-	-	-	-	-	-	-	-
12-14-76	-	-	-	-	-	-	-	-	-	-
12-15-76	8.0	2.9	4.3	-	-	-	-	144	148	26
12-16-76	7.0	3.8	3.8	-	-	-	-	-	-	24
12-17-76	10.0	2.8	3.4	-	-	-	-	-	-	28
12-18-76	-	-	-	-	-	-	-	-	-	-
12-19-76	-	-	-	-	-	-	-	-	-	-
12-20-76	-	-	-	-	-	-	-	-	-	-
12-21-76	-	-	-	-	-	-	-	-	-	-
12-22-76	-	-	-	-	-	-	-	-	-	-
12-23-76	-	-	-	-	-	-	-	-	-	-
12-24-76	-	-	-	-	-	-	-	-	-	-
12-25-76	-	-	-	-	-	-	-	-	-	-
12-26-76	-	-	-	-	-	-	-	-	-	-
12-27-76	13.2	4.4	7.0	-	-	-	-	-	-	-
12-28-76	8.4	3.0	2.4	-	-	-	-	-	-	28
12-29-76	10.0	3.2	3.2	132	148	128	116	216	144	41
12-30-76	-	-	-	-	-	-	-	-	-	-
12-31-76	-	-	-	-	-	-	-	-	-	-
01-01-77	-	-	-	-	-	-	-	-	-	-
01-02-77	9.2	3.1	2.9	-	-	-	-	-	-	50
01-03-77	7.8	1.3	0.8	128	92	144	76	160	164	39
01-04-77	-	0.68	1.00	-	-	-	-	-	-	38
01-05-77	-	1.24	-	-	-	-	-	-	-	-
01-06-77	8.6	1.16	0.88	-	-	-	-	-	-	31
01-07-77	-	-	-	-	-	-	-	-	-	-
01-08-77	-	-	-	-	-	-	-	-	-	-
01-09-77	10.2	1.88	2.12	-	-	-	-	-	-	-
01-10-77	11.1	1.72	1.72	172	172	204	136	216	224	35
01-11-77	8.2	1.64	1.24	-	-	-	-	-	-	43
01-12-77	10.8	0.68	0.88	-	-	-	-	-	-	29
01-13-77	8.4	2.48	2.00	-	-	-	-	-	-	39
01-14-77	-	-	-	-	-	-	-	-	-	-
01-15-77	-	-	-	-	-	-	-	-	-	-
01-16-77	10.2	2.44	2.56	-	-	-	-	-	-	31
01-17-77	10.8	1.68	2.36	216	164	248	168	112	212	35
01-18-77	9.6	1.03	1.16	-	-	-	-	-	-	41
01-19-77	9.1	0.96	0.72	-	-	-	-	-	-	30
01-20-77	8.6	0.50	1.01	-	-	-	-	-	-	36
01-21-77	-	-	-	-	-	-	-	-	-	-
01-22-77	-	-	-	-	-	-	-	-	-	-
01-23-77	9.8	0.56	0.87	-	-	-	-	-	-	34
01-24-77	8.8	1.14	1.27	240	230	270	240	240	220	34
01-25-77	8.1	1.01	1.21	-	-	-	-	-	-	29
01-26-77	8.3	0.72	0.69	-/18	-/18	-/17	-/13	-/46	-/66	34
01-27-77	10.1	0.94	0.64	-	-	-	-	-	-	30
01-28-77	-	-	-	-	-	-	-	-	-	-
01-29-77	-	-	-	-	-	-	-	-	-	-
01-30-77	9.3	0.72	0.88	320/16	200/14	230/19	150/13	300/54	290/73	31

TABLE A-21. FULL SCALE PERFORMANCE DATA
PHOSTRIP LPE SYSTEM - RENO/SPARKS, NEVADA

ORTHO-PHOSPHORUS (mg/1)

DATE	FEED	EFFLUENT		MIXED LIQUOR				AEROBIC RECYCLE				ANOXIC RECYCLE		STRIPPER SUPERNATANT
		1	2	1		2		1		2				
				HIGH	LOW	HIGH	LOW	HIGH	LOW	HIGH	LOW	HIGH	LOW	
11-22-76	5.6	0.60	0.52	0.02	-	0.04	-	3.0	-	5.4	-	116	-	29
11-23-76	5.8	0.60	1.36	0.02	0.03	0.03	0.03	14.0	9.6	12.4	5.0	64	65	23
11-24-76	6.4	0.80	0.52	0.42	0.07	0.54	0.06	12.4	9.0	10.2	5.2	60	76	25
11-25-76	7.6	1.76	1.84	0.07	0.12	0.07	0.17	6.8	7.8	2.8	5.0	60	68	25
11-26-76	7.0	0.82	0.48	0.01	0.07	0.01	0.06	3.8	4.0	3.8	7.8	81	64	24
11-27-76	8.2	1.72	1.36	1.96	1.49	0.82	0.49	9.0	6.6	14.4	6.6	52	64	23
11-28-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-29-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-30-76	7.2	2.88	3.76	1.01	0.07	1.34	0.07	12.2	7.4	9.4	6.0	52	64	27
11-31-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-01-76	6.8	1.96	2.80	1.05	0.09	-	0.13	-	-	-	-	52	64	16
12-02-76	7.4	1.88	2.88	-	0.43	-	0.11	-	-	-	-	-	68	24
12-03-76	6.8	2.16	2.36	-	0.02	-	0.22	-	-	-	-	-	76	16
12-04-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-05-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-06-76	8.0	2.96	2.96	0.59	0.84	0.97	0.97	-	-	-	-	60	88	17
12-07-76	7.6	4.96	5.44	-	-	-	-	-	-	-	-	-	-	37
12-08-76	4.6	2.20	1.50	1.16	0.22	1.52	0.25	-	-	-	-	40	48	25
12-09-76	7.0	2.20	3.50	-	-	-	-	-	-	-	-	-	-	23
12-10-76	7.0	2.50	3.10	1.12	0.07	4.60	0.09	-	-	-	-	44	62	26
12-11-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-12-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-13-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-14-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-15-76	6.0	1.16	3.40	3.16	0.79	3.60	0.36	-	-	-	-	52	56	25
12-16-76	6.2	2.80	2.30	-	-	-	-	-	-	-	-	-	-	9
12-17-76	-	2.40	-	-	-	-	-	-	-	-	-	-	-	26
12-18-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-19-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-20-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-21-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-22-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-23-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-24-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-25-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-26-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-27-76	7.8	3.8	4.4	-	-	-	-	-	-	-	-	-	-	-
12-28-76	5.8	1.8	2.0	0.52	0.03	0.52	0.08	-	-	-	-	52	58	25
12-29-76	7.8	2.6	2.5	0.88	0.09	1.20	0.10	-	-	-	-	50	68	37
12-30-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-31-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-01-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-02-77	6.4	1.7	2.4	-	-	-	-	-	-	-	-	-	-	40
01-03-77	6.6	0.90	0.80	2.12	0.01	0.04	0.03	-	-	-	-	84	78	24
01-04-77	-	0.40	0.70	-	-	-	-	-	-	-	-	-	-	38
01-05-77	7.4	0.56	1.13	0.03	0.03	0.01	0.11	-	-	-	-	70	94	-
01-06-77	6.6	0.80	0.60	-	-	-	-	-	-	-	-	-	-	31
01-07-77	-	-	-	1.48	0.02	0.76	0.03	-	-	-	-	66	-	-
01-08-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-09-77	7.6	1.32	1.36	-	-	-	-	-	-	-	-	-	-	-
01-10-77	6.4	1.36	1.40	0.66	0.03	0.37	0.04	-	-	-	-	78	88	38
01-11-77	6.4	1.60	1.12	-	-	-	-	-	-	-	-	-	-	39
01-12-77	6.6	0.56	0.76	0.02	0.04	0.02	0.05	-	-	-	-	72	78	28
01-13-77	6.8	1.72	1.68	-	-	-	-	-	-	-	-	-	-	42
01-14-77	-	-	-	0.09	0.04	0.08	0.05	-	-	-	-	56	72	-
01-15-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-16-77	7.4	1.84	2.04	-	-	-	-	-	-	-	-	-	-	36
01-17-77	7.0	1.44	1.78	0.05	0.03	0.03	0.03	-	-	-	-	64	76	46
01-18-77	7.2	1.04	1.08	-	-	-	-	-	-	-	-	-	-	42
01-19-77	5.8	0.32	0.60	0.03	0.02	0.02	0.05	-	-	-	-	56	74	34
01-20-77	6.0	0.36	0.96	0.01	0.02	0.03	0.04	-	-	-	-	67	66	43
01-21-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-22-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-23-77	5.7	0.76	0.80	-	-	-	-	-	-	-	-	-	-	30
01-24-77	6.4	0.28	0.76	0.02	0.01	0.01	0.01	-	-	-	-	51	62	37
01-25-77	5.6	0.20	0.44	-	-	-	-	-	-	-	-	-	-	31
01-26-77	6.1	0.28	0.32	0.02	0.01	0.01	0.01	-	-	-	-	59	74	33
01-27-77	6.4	0.44	0.40	-	-	-	-	-	-	-	-	-	-	28
01-28-77	-	-	-	0.04	0.01	0.04	0.01	-	-	-	-	59	68	-
01-29-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-30-77	6.6	0.52	0.52	.03	.04	.05	.05	-	-	-	-	62	82	31

TABLE A-22. FULL SCALE PERFORMANCE DATA
PHOSTRIP LPE SYSTEM - RENO/SPARKS, NEVADA
TSS (mg/l)

DATE	FEED	EFFLUENT 1	EFFLUENT 2	EFFLUENT 3	MIXED LIQUOR 1 HIGH	MIXED LIQUOR 1 LOW	MIXED LIQUOR 2 HIGH	MIXED LIQUOR 2 LOW	MIXED LIQUOR 3 HIGH	MIXED LIQUOR 3 LOW	RECYCLE 1 HIGH	RECYCLE 1 LOW	RECYCLE 2 HIGH	RECYCLE 2 LOW	RECYCLE 3 HIGH	RECYCLE 3 LOW	STRIPPER RETURN HIGH	STRIPPER RETURN LOW	STRIP SUPER
11-22-76	-	16	14	15	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-23-76	-	20	17	17	1,080	1,300	1,040	1,236	1,080	1,082	6,096	3,442	6,216	3,606	2,974	2,344	-	-	-
11-24-76	-	-	-	-	1,190	1,412	1,088	1,294	1,046	1,160	7,318	4,354	5,626	3,196	4,016	6,162	7,730	8,180	-
11-25-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-26-76	-	14	12	7	-	-	-	-	-	-	-	-	-	-	-	-	-	-	150
11-27-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-28-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-29-76	-	20	27	31	862	982	1,144	2,970	970	1,306	3,626	2,566	6,144	3,290	4,602	3,600	-	-	44
11-30-76	-	-	-	-	1,130	1,400	1,012	1,284	1,064	1,126	6,030	3,766	6,220	3,832	4,626	4,666	6,520	-	-
11-31-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-01-76	118	20	21	-	1,006	1,287	1,020	1,206	-	-	6,032	2,822	4,228	3,772	-	-	-	-	44
12-02-76	115	19	19	-	868	780	1,070	1,250	-	-	5,630	2,832	6,394	3,668	-	-	-	-	-
12-03-76	100	22	16	-	1,086	1,274	762	1,224	-	-	8,272	4,072	6,302	3,200	-	-	-	-	-
12-04-76	128	14	19	-	1,000	1,000	1,036	1,036	-	-	3,436	3,436	6,968	4,960	-	-	-	-	-
12-05-76	109	26	32	-	1,132	1,132	1,060	1,060	-	-	5,316	5,316	5,266	5,266	-	-	-	-	62
12-06-76	106	16	19	-	1,116	1,330	1,016	1,230	-	-	6,722	3,022	6,522	3,514	-	-	-	10,610	-
12-07-76	101	23	26	-	1,086	1,170	958	1,196	-	-	7,290	4,054	7,072	4,084	-	-	-	-	-
12-08-76	118	23	28	-	918	1,222	874	1,030	-	-	6,548	3,650	8,966	4,910	-	-	-	-	-
12-09-76	89	7	8	-	1,072	1,278	1,078	1,364	-	-	6,784	3,770	7,736	4,690	-	-	-	-	-
12-10-76	123	20	29	-	952	1,164	1,060	1,250	-	-	7,064	3,568	8,538	4,376	-	-	-	-	-
12-11-76	87	29	30	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-12-76	104	32	29	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-13-76	105	31	28	-	742	1,238	976	1,366	-	-	6,758	3,424	7,884	4,352	-	-	-	-	-
12-14-76	117	18	17	-	970	1,184	1,100	1,344	-	-	6,318	3,376	7,252	4,536	-	-	-	7,760	69
12-15-76	107	23	20	-	968	1,188	1,032	1,330	-	-	6,644	3,096	7,710	4,556	-	-	-	-	65
12-16-76	101	29	21	-	920	1,126	998	1,270	-	-	6,610	3,072	6,616	4,282	-	-	6,080	8,160	-
12-17-76	103	27	26	-	864	1,090	1,042	1,198	-	-	6,266	3,122	7,766	4,184	-	-	-	-	-
12-18-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-19-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-20-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-21-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-22-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-23-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-24-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-25-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-26-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-27-76	113	27	25	-	896	1,238	996	1,334	-	-	6,482	3,464	7,056	4,016	-	-	-	-	-
12-28-76	120	32	29	-	1,028	1,228	1,164	1,478	-	-	6,840	4,020	6,828	4,636	-	-	-	-	70
12-29-76	100	12	8	-	1,276	1,736	1,138	1,570	-	-	7,076	5,320	6,832	5,696	-	-	6,300	9,160	105
12-30-76	119	15	11	-	1,124	1,304	1,108	1,508	-	-	6,708	4,448	7,900	4,584	-	-	-	-	-
12-31-76	91	13	21	-	1,028	1,028	1,234	1,254	-	-	5,580	5,580	6,826	6,826	-	-	-	-	-
01-01-77	86	23	20	-	1,236	1,234	1,366	1,366	-	-	6,062	6,062	6,950	6,950	-	-	-	-	-
01-02-77	113	22	30	-	1,206	1,206	1,372	1,372	-	-	5,772	5,772	5,696	5,696	-	-	-	-	330
01-03-77	122	25	29	-	842	1,360	1,112	1,456	-	-	6,464	4,092	6,912	3,820	-	-	11,950	11,830	-
01-04-77	119	22	18	-	1,074	1,262	1,290	1,736	-	-	6,300	4,054	7,344	4,588	-	-	-	-	70
01-05-77	106	7	10	-	1,118	1,064	1,378	1,388	-	-	6,300	5,644	7,156	6,824	-	-	-	-	-
01-06-77	120	19	22	-	1,138	1,306	1,134	1,222	-	-	7,560	3,928	6,360	5,672	-	-	-	-	-
01-07-77	103	18	17	-	1,138	1,358	1,288	1,492	-	-	7,300	4,996	7,404	4,874	-	-	-	-	-
01-08-77	106	12	16	-	1,160	1,160	1,300	1,300	-	-	6,364	6,364	6,084	6,084	-	-	-	-	-
01-09-77	83	16	17	-	1,118	1,118	1,256	1,256	-	-	5,616	5,616	5,704	5,704	-	-	-	-	-
01-10-77	81	22	19	-	1,064	1,348	1,230	1,434	-	-	7,700	5,712	7,704	5,980	-	-	9,700	10,520	-
01-11-77	91	17	9	-	1,052	1,126	1,092	1,206	-	-	7,752	5,376	5,756	3,988	-	-	-	-	76
01-12-77	100	20	23	-	1,126	1,162	1,096	1,326	-	-	6,784	2,688	6,012	2,764	-	-	-	-	61
01-13-77	98	25	27	-	1,090	1,426	1,174	1,394	-	-	7,300	5,528	6,508	4,344	-	-	-	-	-
01-14-77	91	11	12	-	1,100	1,316	1,176	1,394	-	-	7,276	4,316	6,916	5,104	-	-	-	-	-
01-15-77	109	16	18	-	1,180	1,180	1,278	1,278	-	-	6,776	6,776	5,952	5,952	-	-	-	-	-
01-16-77	111	20	24	-	1,110	1,110	1,192	1,192	-	-	6,186	6,186	6,336	6,336	-	-	-	-	57
01-17-77	107	20	18	-	1,114	1,302	1,074	1,396	-	-	7,364	4,736	7,268	5,000	-	-	8,250	8,560	-
01-18-77	102	19	17	-	1,050	1,254	1,076	1,400	-	-	7,300	5,284	6,932	5,290	-	-	-	-	-
01-19-77	96	21	24	-	1,114	1,232	1,062	1,216	-	-	6,660	6,560	5,792	4,980	-	-	-	-	56
01-20-77	99	18	16	-	1,066	1,268	1,070	1,296	-	-	6,452	6,620	7,556	5,703	-	-	-	-	-
01-21-77	93	38	30	-	1,232	1,174	1,106	1,036	-	-	7,334	4,232	6,640	6,192	-	-	-	-	-
01-22-77	84	32	40	-	1,206	1,206	1,192	1,192	-	-	6,664	6,664	6,464	6,464	-	-	-	-	-
01-23-77	110	16	21	-	1,236	1,236	1,324	1,324	-	-	6,480	6,480	6,044	6,044	-	-	-	-	36
01-24-77	106	20	23	-	1,038	1,256	1,106	1,356	-	-	7,376	6,544	6,376	6,560	-	-	8,150	9,060	-
01-25-77	113	27	25	-	1,018	1,288	1,180	1,360	-	-	6,828	6,272	8,100	6,748	-	-	-	-	65
01-26-77	116	18	26	-	1,026	1,394	1,106	1,368	-	-	3,376	4,400	8,160	4,980	-	-	-	-	-
01-27-77	106	20	23	-	1,142	1,332	1,184	1,256	-	-	7,313	6,352	7,030	4,988	-	-	-	-	-
01-28-77	91	15	18	-	1,062	1,076	1,098	1,258	-	-	7,936	5,860	7,712	4,492	-	-	-	-	-
01-29-77	99	10	13	-	1,110	1,110	1,262	1,262	-	-	6,064	6,064	6,796	6,796	-	-	-	-	-
01-30-77	113	21	21	-	1,128	..128	1,170	1,170	-	-	6,656	6,656	6,648	6,648	-	-	-	-	67

TABLE A-23. FULL SCALE PERFORMANCE DATA
PHOSTRIP LPE SYSTEM - RENO/SPARKS, NEVADA
VSS (mg/1)

DATE	FEED	EFFLUENT			MIXED LIQUOR						RECYCLE 1		RECYCLE 2		RECYCLE 3		STRIPPER RETURN		STRIP SUP
					1		2		3										
		1	2	3	HIGH	LOW	HIGH	LOW	HIGH	LOW	HIGH	LOW	HIGH	LOW	HIGH	LOW	HIGH	LOW	
11-22-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-23-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-24-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-25-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-26-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-27-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-28-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-29-76	-	-	-	-	650	704	902	2,272	802	1,058	2,878	2,032	4,790	4,116	3,736	2,790	-	-	-
11-30-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
11-31-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-01-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-02-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-03-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-04-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-05-76	30	23	28	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-06-76	-	-	-	-	922	1,120	844	1,016	-	-	-	-	-	-	-	-	-	-	53
12-07-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-08-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	49
12-09-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-10-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-11-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-12-76	100	28	25	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-13-76	-	-	-	-	774	1,036	794	1,114	-	-	-	-	-	-	-	-	-	-	-
12-14-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	46
12-15-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	47
12-16-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-17-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-18-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-19-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-20-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-21-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-22-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-23-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-24-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-25-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-26-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-27-76	88	14	17	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12-28-76	-	-	-	-	812	966	952	1,206	-	-	5,560	3,260	5,592	3,792	-	-	-	-	-
12-29-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	79
12-30-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	99
12-31-76	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-01-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-02-77	90	16	23	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-03-77	-	-	-	-	681	1,000	911	1,144	-	-	-	-	-	-	-	-	-	-	296
01-04-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-05-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-06-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-07-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-08-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-09-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-10-77	-	-	-	-	884	1,072	994	1,146	-	-	5,988	4,456	5,568	4,724	-	-	-	-	-
01-11-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-12-77	79	20	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-13-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	60
01-14-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-15-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-16-77	92	12	17	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-17-77	-	-	-	-	888	1,022	856	1,112	-	-	5,864	3,728	5,688	3,904	-	-	6,780	7,280	50
01-18-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-19-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-20-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	56
01-21-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-22-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-23-77	107	-	21	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-24-77	-	-	-	-	840	1,018	890	1,090	-	-	5,820	5,164	5,052	5,164	-	-	-	-	50
01-25-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-26-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-27-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-28-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-29-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
01-30-77	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-